U.S. DEPARTMENT OF COMMERCE
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PB80-152812

Benzene Emissions from Maleic Anhydride Industry - Background Information for Proposed Standards

(U.S.) Environmental Protection Agency, Research Triangle Park, NC

Feb 80

United States
Environmental Protection
Agency

Office of Air Quality
Planning and Standards
Research Triangle Park NC 27711

EPA-450/3-80-001a
February 1980

Air

PB80-152812

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Benzene Emissions from Maleic Anhydride Industry — Background Information for Proposed Standards

Emission Standards and Engineering Division

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air, Noise, and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

February 1980

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PUBLICATION NO. EPA-450/3-80-001a

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-450/3-80-001a	2.	3. RECIPIENT'S ACCESSION NO. PB80-152812
4. TITLE AND SUBTITLE Benzene Emissions from the Maleic Anhydride Industry — Background Information for Proposed Standards		5. REPORT DATE February 1980 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Office of Air Quality Planning and Standards Environmental Protection Agency Research Triangle Park, North Carolina 27711		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS DAA for Air Quality Planning and Standards Office of Air, Noise, and Radiation U.S. Environmental Protection Agency Research Triangle Park, North Carolina 27711		13. TYPE OF REPORT AND PERIOD COVERED Final 14. SPONSORING AGENCY CODE EPA/200/04

15. SUPPLEMENTARY NOTES

16. ABSTRACT

A National Emission Standard for the control of benzene emissions from maleic anhydride plants is being proposed under the authority of section 112 of the Clean Air Act. The proposed standard would apply to both new and existing sources. This document contains background information and environmental and economic assessments of the regulatory alternatives considered in developing the proposed standard.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Pollution Control National Emission Standards for Hazardous Air Pollutants Maleic Anhydride Plants Benzene Hazardous Pollutants	Air Pollution Control	13b

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)

Background Information
and Draft
Environmental Impact Statement
for
Benzene Emissions from Maleic Anhydride Plants

Type of Action: Administrative

Prepared by:

[illegible] (Date)

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Approved by:

1/22/80 (Date)

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Environmental Protection Agency
Washington, D.C. 20460

Draft Statement Submitted to EPA's
Office of Federal Activities for Review on 2/80 (Date)

This document may be reviewed at:

Central Docket Section
Room 2903B, Waterside Mall
Environmental Protection Agency
401 M Street, S.W.
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INTRODUCTION

This document describes the domestic maleic anhydride production industry with emphasis on the benzene emissions to the atmosphere. The first sections of the report discuss the characteristics of the industry, the processes used to produce maleic anhydride, and the associated emissions. Only ducted process vents are covered in this document. Fugitive emission sources of benzene, storage and handling sources of benzene, and miscellaneous secondary sources of benzene will be covered in other documents in preparation. Applicable control techniques and alternative levels of control and their environmental, energy, cost, and economic impacts are described.

SUMMARY

REGULATORY ALTERNATIVES

Two regulatory alternatives were considered for best available technology (considering environmental, energy, and economic impacts) (BAT) to control benzene emissions from existing process vents of plants that use benzene as a feedstock to manufacture maleic anhydride or maleic acid. These alternatives are:

- A 97-percent reduction from uncontrolled emission levels, based on the best control that has been achieved at an existing maleic plant using universally applicable equipment; and
- A 99-percent reduction from uncontrolled levels, based on the best control considered feasible using technology transfer.

While no particular control technology would be specifically required for either alternative, carbon adsorption or thermal incineration could be used in achieving either 97 or 99 percent control.

A third alternative, a 100-percent reduction in benzene emissions based on the use of a substitute feedstock, such as n-butane, was considered as an alternative for BAT for new sources. This alternative was not considered as an alternative for BAT for existing sources because of uncertainties about the technical feasibility of converting each existing source to n-butane and the impacts of such a conversion. However, few new sources are expected to be built until the mid-1980's. Thus, a new source could be designed to use n-butane and would therefore not encounter the potential problems associated with conversion.

ENVIRONMENTAL IMPACTS--EXISTING SOURCES

Table S-1 summarizes the environmental and economic impacts of the regulatory alternatives considered. These alternatives would reduce nationwide benzene emissions from maleic anhydride plants from 6,400 Mg/yr to 1,030 Mg/yr for 97 percent control, to 540 Mg/yr for 99 percent control,

TABLE S-1. ENVIRONMENTAL AND ECONOMIC IMPACTS OF REGULATORY ALTERNATIVES

Alternative	Impact on benzene	Impact on HC	Other air impacts	Water impact	Solid waste impact	Energy impact	Space impact	Noise impact	Economic impact
97% control	+3	+3	-1	-1	-1	-1	-1	-1	-2
99% control	+4	+4	-1	-1	-1	-3	-1	-1	-3
Substitution of feedstock, such as n-butane (new plants only)	+5	-2	-1	-1	0	0	0	0	+2
Delayed standard	-2	-2							0
No standard	-3	-3							0

KEY			
+	Beneficial impact	0	No impact
-	Adverse impact	1	Negligible impact
		2	Small impact
		3	Moderate impact
		4	Large impact
		5.	Larger impact

and to 0 Mg/yr for 100 percent control. These emissions include fugitive, storage, and handling emissions, which are not covered by the proposed standard.

The control systems likely to be used to meet a 97-percent or 99-percent benzene emission reduction (incineration or carbon adsorption) would also reduce emissions of other hydrocarbons, which might be toxic and which contribute to photochemical oxidant formation and associated environmental problems.

The reduction in nationwide benzene emissions would result in minimal, adverse environmental impacts. These adverse impacts would include small increases in nitrogen oxide because of high incineration temperatures and sulfur oxide emissions into the air, if fuel oil were used as a supplemental fuel for incinerators. There would be small increases in solid wastes and in benzene in wastewater. Compliance with the alternatives would increase nationwide energy consumption by an estimated 50,000 barrels for 97 percent control and 85,000 barrels of fuel oil equivalent per year for 99 percent control by 1980.

ECONOMIC IMPACTS--EXISTING SOURCES

The capital investment required by the domestic maleic anhydride industry to comply with the proposed standard would be about $6.6 million for 97 percent control and $9.1 million for 99 percent control over the 2-year period from 1979 to 1981. The total annualized operating costs of the industry would increase by about $2.5 million/yr for 97 percent control and $4.5 million for 99 percent control by 1983. If continuous monitoring were required, annualized costs would be increased by an additional $9,000 for each plant. The market price of maleic anhydride would increase an estimated 1.2 percent for 97 percent control and 1.7 percent for 99 percent control. In addition, one plant might close if the standard were based on 97 percent control and two plants might close if the standard were based on 99 percent control.

ENVIRONMENTAL IMPACTS--NEW SOURCES

The use of a substitute feedstock by new sources would eliminate benzene emissions from process vents, storage and handling of benzene, and fugitive sources. Information indicates there would be greater quantities of volatile organic compounds (VOC's) from the uncontrolled n-butane process than from the uncontrolled benzene process. EPA is currently developing a new source performance standard for air oxidation processes in the chemical industry. This standard would cover new sources producing maleic anhydride from n-butane.

ECONOMIC IMPACTS--NEW SOURCES

A plant using the n-butane process has been estimated to achieve as much as a 7.3¢/kg (3.3¢/lb) maleic anhydride cost reduction below benzene-based maleic anhydride because of the difference in feedstock costs. This reduction is achieved despite a greater annualized utilities cost recently estimated to be \$450,000 for the n-butane-based process compared to \$350,000 for the benzene-based process of the same size. Capital costs for the n-butane process may be greater because of the larger reactors and higher temperature required. Because of the cost reduction, new plants are expected to use the n-butane process regardless of EPA's requirement.

1. THE MALEIC ANHYDRIDE INDUSTRY

1.1 GENERAL

This chapter discusses maleic anhydride production, focusing primarily on three basic processes used in the United States and the benzene emissions related to these processes. The processes are: oxidation of benzene, oxidation of n-butane, and recovery as a byproduct of phthalic anhydride manufacture. Most of the current domestic capacity is based on the oxidation of benzene. Foreign processes are also briefly described.

Table 1-1 shows the end uses of maleic anhydride and their expected growth rates. The predominant end use is the production of unsaturated polyester resins, which go into reinforced plastic applications such as marine craft, building panels, automobiles, tanks, and pipes. The United States maleic anhydride production capacity for 1977 was reported to be 229,000 Mg (5.0×10^8 lb) with only 56 percent of this capacity currently used.[1,2] At an estimated 11-percent annual growth in maleic anhydride consumption, production would reach 95 percent of present capacity by 1982. No shortage of benzene, the major raw material, is expected during this period.[2]

As of 1977, there were eight producers of maleic anhydride in the United States with 10 plants. However, in 1979, Koppers announced the mothballing of its Pennsylvania facility.[3] Table 1-2 lists the producers and the processes being used, while Figure 1-1 shows the plant locations. Approximately 82 percent of the 229,000-Mg/yr (5.0×10^8 lb/yr) domestic capacity is based on oxidation of benzene; 16 percent capacity comes from oxidation of n-butane; and the remaining 2 percent results from phthalic anhydride production, which yields maleic anhydride as a byproduct.[1] Because of anticipated increases in the price of benzene, work began in 1960 to develop a catalyst suitable for producing maleic anhydride from

TABLE 1-1. MALEIC ANHYDRIDE USAGE AND GROWTH

End use	1978 demand Gg/yr (in 1,000 lb/yr)	1978 demand as % of production	Average annual % growth 1978-83
Unsaturated polyester resins	77.7 (171,200)	58	13
Agricultural chemicals	13.4 (29,600)	10	10
Lubricating additives	8.7 (19,200)	7	12
Fumaric acid	6.7 (14,800)	5	5
Copolymers	5.9 (13,000)	4	9
Maleic acid	4.2 (9,200)	3	10
Reactive plasticizers	4.0 (8,800)	3	8
Surface-active agents	3.2 (7,000)	2	8
Alkyd resins	1.5 (3,400)	1	4
Chlorendic anhydride and acid	1.2 (2,600)	1	13
Other	7.5 (16,600)	6	5
All MA products	134.0 (295,400)	100	11

SOURCES: Chemical Profile on Maleic Anhydride. Chemical Marketing Reporter. February 18, 1978.

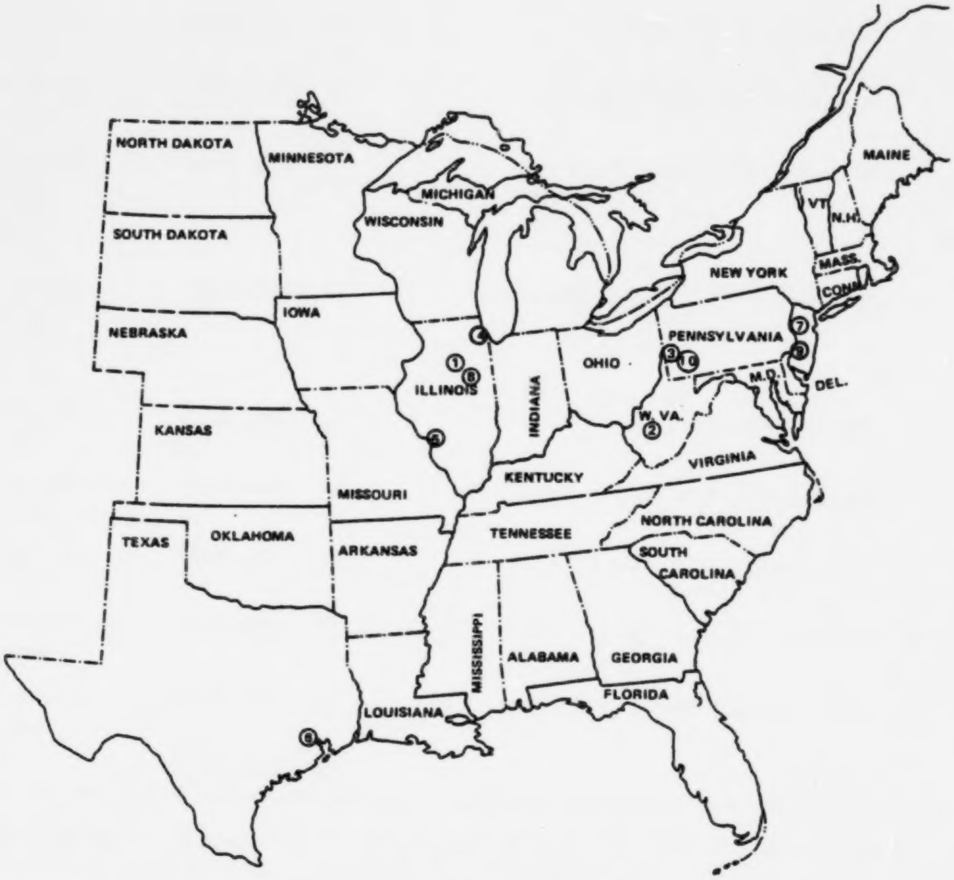
Blackford, J. C. Marketing Research Report on Maleic Anhydride. In: Chemical Economics Handbook. Menlo Park, Stanford Research Inst. July 1976.

TABLE 1-2. MALEIC ANHYDRIDE CAPACITY

		Capacity--1977 10^3 Mg (lb)	Process
1.	Amoco, Joliet, Ill.	27 (59,600)	Oxidation of n-butane
2.	Ashland, Neal, W.Va.	27 (59,600)	Oxidation of benzene
3.	Koppers, Bridgeville, Pa.	15 (33,000)	Oxidation of benzene
4.	Koppers, Chicago, Ill.	5 (11,000)	Byproduct of phthalic anhydride manufacture
5.	Monsanto, St. Louis, Mo.	48 (105,800)	(80%) oxidation of benzene (20%) oxidation of n-butane
6.	DENKA, Houston, Tex.	23 (50,800)	Oxidation of benzene
7.	Reichhold, Elizabeth, N.J.	14 (30,800)	Oxidation of benzene
8.	Reichhold, Morris, Ill.	20 (44,000)	Oxidation of benzene
9.	Tenneco, Fords, N.J.	12 (26,400)	Oxidation of benzene
10.	U.S. Steel, Neville Island, Pa.	38 (83,800)	Oxidation of benzene
	TOTAL	229 (504,800)	

SOURCE: Blackford, J. C. Marketing Research Report on Maleic Anhydride. In: Chemical Economics Handbook. Menlo Park, Stanford Research Inst. July 1976.

Letter from He[illegible]tt, P. S., Reichhold Chemicals, Inc., to Patrick, D. R., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 27, 1978.


NORTH DAKOTA
MINNESOTA
MICHIGAN
WISCONSIN
MAINE
VT
N.H.
MASS.
CONN
SOUTH DAKOTA
NEW YORK
IOWA
NEBRASKA
PENNSYLVANIA
OHIO
ILLINOIS
INDIANA
M.D.
DEL.
W. VA.
KANSAS
VIRGINIA
KENTUCKY
MISSOURI
NORTH CAROLINA
TEXAS
OKLAHOMA
ARKANSAS
TENNESSEE
SOUTH CAROLINA
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
LOUISIANA
1
2
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4
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10

KEY

1. Amoco, Joliet, Ill.
2. Ashland, Neal, W. Va.
3. Koppers, Bridgeville, Pa.
4. Koppers, Chicago, Ill.
5. Monsanto, St. Louis, Mo.
6. DENKA, Houston, Tex.
7. Reichhold, Elizabeth, N.J.
8. Riechhold, Morris, Ill.
9. Tenneco, Fords, N.J.
10. U.S. Steel, Neville Island, Pa.

Figure 1-1. Manufacturing locations of maleic anhydride.

SOURCE: Hydroscience. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry: Maleic Anhydride Product Report. March 1978.

n-butane/butene (C_4) streams available from naphtha cracking. This effort was curtailed during the 1961-1967 period when the maleic anhydride market was depressed and low-cost benzene was available. In 1967, however, demand for maleic anhydride increased, and Kasei Mizuishima renewed work in Japan. In 1974, DENKA, Chem Systems, BASF, Bayer, Alusuisse/UCB, and Mitsubishi made announcements concerning the production of maleic anhydride from C_4 feedstocks.[4] Presently, Amoco and Monsanto are producing maleic anhydride from an n-butane feedstock,[1] and Halcon Catalyst Industries has completed a plant to produce a catalyst to convert n-butane to maleic anhydride.[5]

In 1979 Monsanto announced plans to build a 45,000-Mg/yr (100×10^6 lb/yr) maleic anhydride plant at Pensacola, Florida, using a proprietary n-butane process.[6] The plant is scheduled for completion in 1983. Later in 1979, DENKA and Badger Company announced plans to build a demonstration plant at DENKA's facility in Houston that would use n-butane as a feedstock.[7] The projected growth rate for the n-butane oxidation process through 1982 is 24.3 percent, compared to only a 9.1-percent growth rate for the benzene oxidation process. This assumes the n-butane process becomes fully commercialized. Substantive data regarding the economic incentives for switching to n-butane oxidation were not available when this document was prepared. Finally, no growth in the quantity of maleic anhydride recovered during phthalic anhydride production is expected.[2]

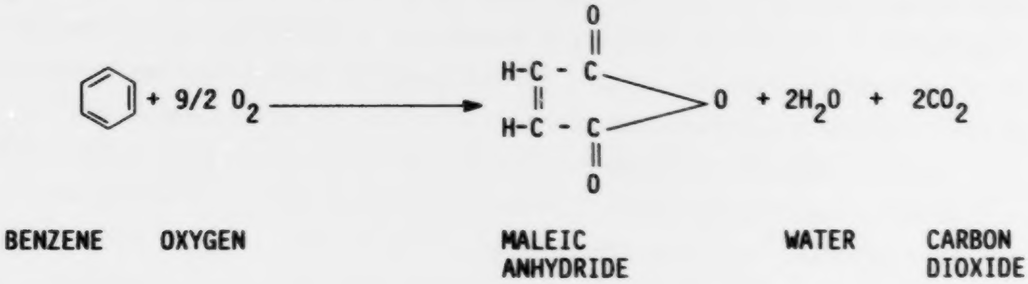
1.2 PROCESS DESCRIPTIONS AND EMISSIONS

The two major processes used to manufacture maleic anhydride in the United States are benzene oxidation and n-butane oxidation, while a small amount of maleic anhydride is recovered as a byproduct of phthalic anhydride production. The only significant foreign process for maleic anhydride production not used in the United States starts with a butene mixture feedstock. While this process is currently used in France,[8] no plans are known to introduce this process domestically.

This chapter addresses only ducted process emissions (i.e., normal vents from the process equipment) and does not include emissions from the storage and handling of benzene, leaks from equipment, or secondary sources of benzene (e.g., benzene evaporation from a wastewater stream). It is important to recognize that no benzene emissions result from the recovery process from phthalic anhydride production. No benzene emissions are anticipated from any process using n-butane or butenes, and none has been found at lower detection limits of 1 ppmv.[9]

1.2.1 Benzene Oxidation Process

Maleic anhydride is produced by the following vapor phase chemical reaction:



The process flow diagram with numbered streams shown in Figure 1-2 represents a typical process. This typical process is continuous, although some plants operate dehydration and distillation batchwise. The emissions in either case are judged to be equivalent.[1]

A mixture of benzene and air enters a tubular reactor where the catalytic oxidation of benzene is carried out at a temperature of 350° to 400° C (662° to 752° F). The catalyst contains approximately 70 percent vanadium pentoxide supported on an inert carrier; most of these catalysts also contain 25 to 30 percent molybdenum oxide. The reaction is highly exothermic, releasing 24.4 MJ/kg (10,514 Btu/lb) of reacted benzene, with the excess heat being used to generate steam. Maleic anhydride yields range from 60 to 67 percent of theoretical.[1]

The reactor feed mixture is provided with excess air to keep the benzene concentration below its lower explosive limit of 1.5 volume percent. The resultant large volume of reactor exhaust (Stream 3) directly influences the size of the subsequent product recovery equipment. After reaction, the stream passes through a cooler, partial condenser, and separator in which 40 percent of the maleic anhydride is condensed and separated as crude product (Stream 4).[10] The remaining product and other organics (Stream 5) enter the product recovery absorber, where they contact water or aqueous maleic acid. The liquid effluent from the absorber (Stream 6) is a 40 percent (by weight) aqueous solution of maleic acid. The absorber vent

(Vent A) exhausts to the atmosphere or is directed to an emission control device.[1]

The 40-percent maleic acid (Stream 6) is dehydrated by azeotropic distillation with xylene. Any xylene retained in the crude maleic anhydride (Stream 9) is removed in a xylene-stripping column, and the crude maleic anhydride (Stream 10) from this column is then combined with the crude maleic anhydride from the separator (Stream 4).[1]

Color-forming impurities in the crude maleic anhydride are removed by aging, which causes impurities to polymerize. After aging, the crude maleic anhydride (Stream 11) is fed to a fractionation column that yields purified molten maleic anhydride as the overhead product (Stream 12). The fractionation column bottoms containing the color-forming impurities are removed as liquid residue waste (Stream 13). This stream is either combined with other effluent or is fed to a liquid incinerator.[1] A small percentage of the finished product is made into briquettes.

The vacuum lines from the dehydration column, xylene stripper, and fractionation column are joined to the vacuum system (Stream 14). The refining vacuum system vent (Vent B) can exhaust to the atmosphere, recycle to the product recovery absorber (Stream 5), or be directed to an emission control device. Water from the vacuum system can be recycled as makeup water (Stream 7) or join the liquid residue waste (Stream 13).[1]

Essentially, all process emissions exit through the product recovery absorber (Vent A). These emissions include any unreacted benzene; this is typically 3 to 10 percent of the total benzene feed.[11] The only other process emission source is the refining vacuum system vent (Vent B), which may contain small amounts of maleic anhydride, xylene, and a slight amount of benzene, because benzene could be absorbed in the liquid stream from the product recovery absorber (Stream 6) or in the crude maleic anhydride from the separator (Stream 4).

Analyses by producers have detected no benzene at the 10-ppm level in the final maleic anhydride products produced by the benzene process.[12,13,14]

Although the process just described and illustrated in Figure 1-2 typifies the benzene oxidation process, variations exist. In place of the partial condensation system (cooler, partial condenser, and separator)

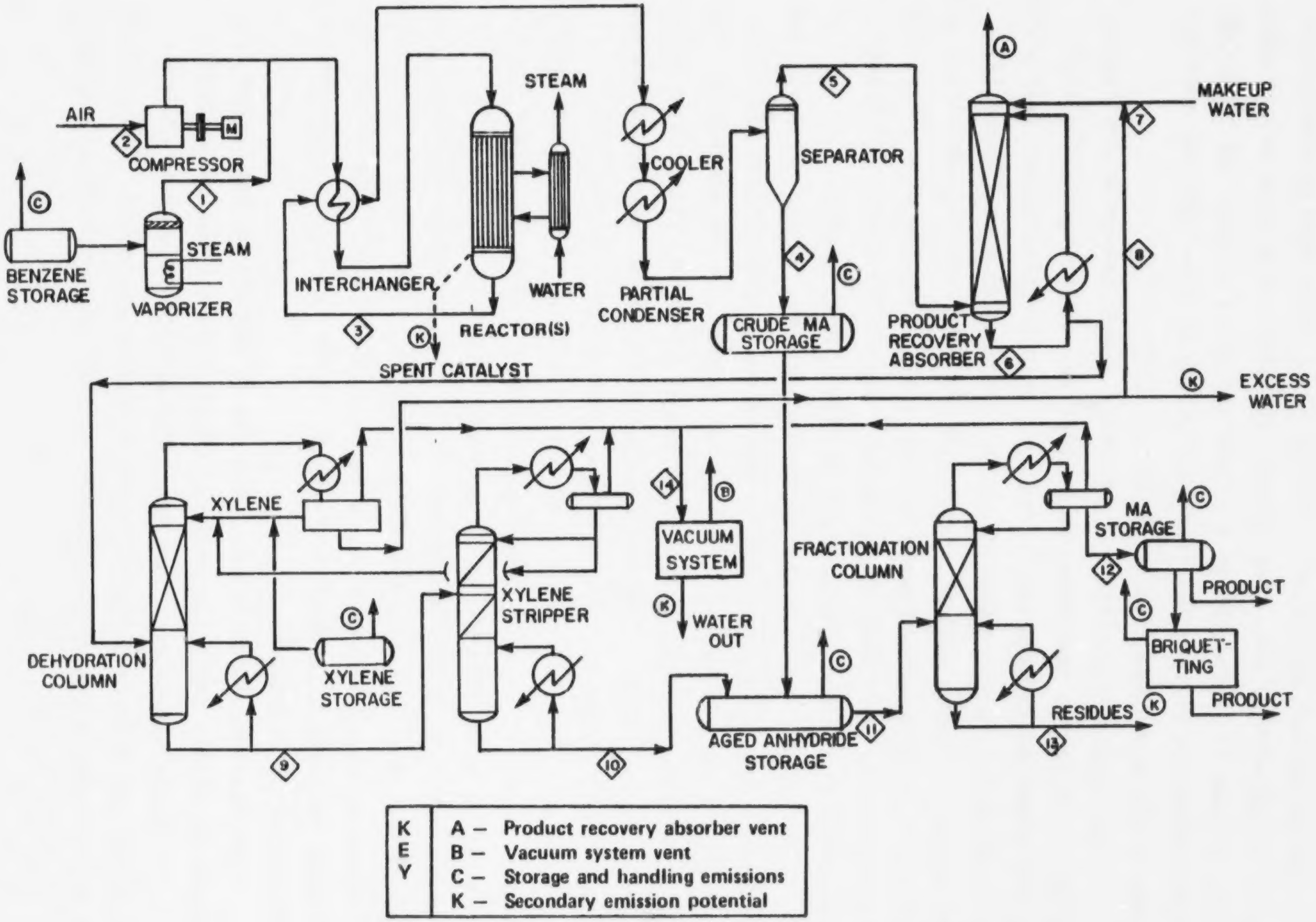

AIR
COMPRESSOR
M
BENZENE STORAGE
VAPORIZER
STEAM
INTERCHANGER
REACTOR(S)
STEAM
WATER
SPENT CATALYST
COOLER
PARTIAL CONDENSER
SEPARATOR
CRUDE MA STORAGE
PRODUCT RECOVERY ABSORBER
MAKEUP WATER
EXCESS WATER
DEHYDRATION COLUMN
XYLENE
XYLENE STORAGE
XYLENE STRIPPER
VACUUM SYSTEM
WATER OUT
AGED ANHYDRIDE STORAGE
FRACTIONATION COLUMN
MA STORAGE
PRODUCT
BRIQUETTING
PRODUCT
RESIDUES
1
2
3
4
5
6
7
8
9
10
11
12
13
14
A
B
C
K
KEY
A — Product recovery absorber vent
B — Vacuum system vent
C — Storage and handling emissions
K — Secondary emission potential

Figure 1-2. Process flow diagram for uncontrolled model plant.

shown in Figure 1-2, a so-called switch condensing system can be incorporated. This system uses a series of condensers that are alternately cooled to freeze maleic anhydride on the surface and heated to melt the maleic anhydride for pumping to crude maleic anhydride storage. Switch condensing can remove up to 60 percent of the maleic anhydride from the process compared to 40 percent for the partial condensation system.[15] Removal of this additional maleic anhydride would allow the size of the product recovery absorber to be reduced and would slightly reduce the maleic acid loss through the product recovery absorber (Vent A).

Xylene is the only known azeotropic agent currently used for dehydration, although several other agents can be used, including isoamyl butyrate, di-isobutyl ketone, anisole, and cumene.[1] A vacuum evaporation system, which replaces the dehydration column and xylene stripper, is used by at least one plant to remove water and dehydrate the maleic acid to form maleic anhydride.[15] Because an azeotropic agent is not required in that case, xylene is eliminated as a process emission.

The emission rates for the benzene oxidation process are based on a model plant with a capacity of 22,700 Mg/yr (50×10^6 lb/yr), assuming 8,000 hours of annual operation and 94.5 percent conversion of benzene. Though not an actual plant, it is typical of most plants in capacity and operating range. The model benzene oxidation process, shown in Figure 1-2, best fits today's maleic anhydride manufacturing and engineering technology. Single-process trains as shown are typical for the large plants except in the reaction area where multiple reactors are common. The model process uses partial condensation and azeotropic distillation with xylene. The emission rates and sources for the benzene oxidation process are summarized in Table 1-3 and represent compilation of data from several plants.

Emission rates depend on conversion. Conversion is the ratio of the amounts of raw material reacted to the amount of raw material fed to the reactor--in this case, benzene reacted to benzene fed. If conversion were 100 percent, there would be no benzene emissions from the process vent because all of the benzene would be converted to something else. The exact conversion rate is important. A maleic anhydride plant operating at a 95-percent benzene conversion would require a control system 80 percent efficient in order to emit the same amount of benzene as an identical, uncontrolled plant operating at 99 percent benzene conversion.

TABLE 1-3. BENZENE AND TOTAL VOLATILE ORGANIC COMPOUND (VOC) EMISSIONS FROM MODEL UNCONTROLLED MALEIC ANHYDRIDE PLANT[a]

Source	Stream designation Figure 1-2	kg (lb) of benzene per Mg (1,000 lb) of MA produced	Total kg (lb) of VOC per Mg (1,000 lb) of MA produced	Benzene emitted kg/hr (lb/hr)	Total VOC emitted kg/hr (lb/hr)
Product recovery absorber	A	67.0 (67.0)	86.0 (86.0)	190 (418)	244 (537)
Refining vacuum system	B	---	0.1 (0.1)	---	0.28 (0.62)

[a] Emission rates are annual averages based on 8,000 hr/yr of operation, 22,700-Mg/yr (50×10^6 lb/yr) capacity, 94.5% conversion.

SOURCES: Morse, P. L. Maleic Anhydride. In: Process Economic Program. Menlo Park, Stanford Research Inst. November 1973.

Lawson, J. F. Trip Report for Visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977. Hydroscience, Inc. EPA Contract Number 68-02-2577.

Pervier, J. W., et al. Survey Reports on Atmospheric Emissions from the Petrochemical Industry, Volume III. Houdry Division of Air Products, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA-450/3-73-005-c. April 1974.

The highest possible benzene conversion is not necessarily optimum because not all the benzene reacted forms maleic anhydride. In fact, at high conversion rates, the yield (ratio of product formed to reactant consumed) is often lower than at lower conversions. Thus, production often increases with operation at a lower conversion; i.e., that which best balances conversion and yield.

The composition of the waste stream from the product recovery absorber--the location of the main process vent--of a model plant is shown in Table 1-4. All plants have this vent. The benzene-to-air ratio influences the concentrations of emissions from this vent. Excess air must be fed to the reactor to maintain the benzene concentration below its lower explosive limit.

Some types of process upsets result in more benzene being released because the product recovery absorber can only remove benzene up to its solubility level in water or aqueous maleic acid. These upsets can result in short-duration benzene and volatile organic compound (VOC) emissions of three to five times normal. Process startup at an uncontrolled plant also results in temporary benzene emissions three to five times the normal rate because the benzene does not react completely until proper catalyst temperature is reached and, again, absorption is only effective up to the limits of solubility.

The refining vacuum system vent (Vent B, Figure 2) exhausts the noncondensibles from the three vacuum columns used to dehydrate and fractionate maleic anhydride. The VOC emissions--maleic acid, xylene, and possibly benzene--are estimated to be relatively small, as indicated in Table 1-3. Process upsets, startups, and shutdowns do not affect the VOC emissions from this vent.[16]

Table 1-5 summarizes the estimated benzene emissions from maleic anhydride plants for uncontrolled and existing control conditions.

Descriptions of fugitive and storage benzene emissions are to be covered in another document.

1.2.2 n-Butane Oxidation Process

Little information in the open literature is available on maleic anhydride production by the oxidation of n-butane, as currently practiced in the United States. Thus, it is particularly difficult to assess the effi-

TABLE 1-4. WASTE GAS COMPOSITION--PRODUCT RECOVERY ABSORBER

Component	Weight % (weighted average)	kg/hr	(lb/hr)
Oxygen	16.67	13,800	(30,360)
Nitrogen	73.37	60,740	(133,628)
Carbon dioxide	3.33	2,757	(6,065)
Carbon monoxide	2.33	1,929	(4,244)
Water	4.00	3,312	(7,286)
Benzene	0.23	190	(418)
Maleic acid	0.01	8	(18)
Formaldehyde	0.05	41	(90)
Formic acid	0.01	8	(18)
		82,785	(182,127)

SOURCE: Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

TABLE 1-5. BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS[1]

	Capacity (Mg/yr)	Total uncontrolled emissions[2] (Mg/yr)	Existing control device efficiency (%)	Emissions with existing control[3] (Mg/yr)
Ashland, Neal W. Va.	27,200	1,849	0	1,849
Koppers, Bridgeville, Pa.	15,400	1,047	99	26
Monsanto, St. Louis, Mo.	38,100	2,589	0	2,589
DENKA, Houston, Tex.	22,700	1,543	97	68
Reichhold, Elizabeth, N.J.	13,600	924	97	44
Reichhold, Morris, Ill.	20,000	1,359	90	154
Tenneco Fords, N.J.	11,800	802	0	802
U.S. Steel, Neville Island, Pa.	38,500	2,616	90	295
Totals		12,729		5,824

[1] Emission rates are based on a benzene conversion rate of 94.5 percent and on: Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

[2] Total uncontrolled emissions

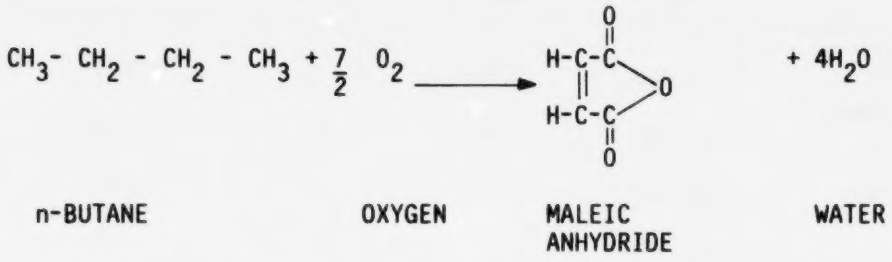
$$= [(2.6 \text{ kg/hr} \times 8{,}760 \text{ hr}) + (190 \text{ kg/hr} \times 8{,}000 \text{ hr})] \times \frac{\text{production capacity Mg/yr}}{\text{model plant capacity } 22{,}700 \text{ Mg/yr}} .$$

[3] Emissions with existing control

$$= \{(2.6 \text{ kg/hr} \times 8{,}760 \text{ hr}) + [190 \text{ kg/hr} \times (1\text{-control device efficiency}) \times 8{,}000 \text{ hr}]\} \times \frac{\text{production capacity Mg/yr}}{\text{model plant capacity } 22{,}700 \text{ Mg/yr}} .$$

ciency of this process relative to the conventional benzene oxidation process. General information on n-butane oxidation is summarized below, although it may not accurately represent the process as practiced by Monsanto and Amoco.

The overall chemical reaction of interest is:



The reaction involves both dehydrogenation and oxidation. The actual catalysts used today have not been reported, although several possibilities are listed in Hydrocarbon Processing. One is a mixture of iron and vanadium pentoxide-phosphorous pentoxide on a silica-alumina support with unspecified proportions. The reaction proceeds at 500° C (932° F) at atmospheric pressure and results in a yield of 14.2 mole percent. Another scheme uses a cerium chloride, cobalt-molybdenum oxide catalyst supported on silica. The cerium chloride dehydrogenates the n-butane to butene, after which the cobalt-molybdenum oxide catalyst oxidizes the butene to maleic anhydride. The reaction is carried out at 490° C (914° F), and a yield of 63 weight percent is reported. A third catalyst system has been reported using phosphorous and vanadium with yields of 55 to 60 percent.[17] The actual catalyst used by the two domestic producers (Monsanto and Amoco) of maleic anhydride from n-butane has not been reported. Catalyst development research is actively continuing, and the catalyst technology is a closely held secret.

The DENKA plant was originally designed to use the cis- and trans-isomers of 2-butene as a feedstock. This feedstock was used for 4 years before being replaced by the more economical benzene process. Later, DENKA experimented with n-butane for the life of one catalyst--6 to 9 months--and encountered problems with catalyst stability.[18]

In 1979, DENKA and the Badger Company announced plans to build and operate a demonstration plant at DENKA's facility to use n-butane. The process uses a fluidized bed reactor.[7] Badger reports high yields and low utilities consumption, based on laboratory and model plant tests.

A simplified flow diagram from one producer indicates that the n-butane oxidation process is similar to the benzene oxidation process shown previously in Figure 1-2. One obvious, major difference is in the raw material storage facilities, where n-butane requires additional safety and property protection safeguards.[19] Because n-butane is in the gaseous state at standard conditions, it will normally be stored as a liquid in pressurized vessels, resulting in additional operating costs. Like the benzene process, product recovery is by partial condensation, product absorption, dehydration, and fractionation.[20] Further indirect evidence supports the assumption that the n-butane process is analogous to the benzene process. First, Ashland Chemical, which started up a new benzene-based plant in 1976, states it is able to convert to n-butane feed simply with a change in catalyst.[21] Second, it has been reported that the switch to n-butane feed can be accomplished at a relatively small fraction of the cost of a new plant.[21] Third, the benzene oxidation process can be converted to n-butane oxidation by changing the catalyst system, and conversion can be accomplished for much less than the cost of a new plant.[21] This implies that, for the most part, the same unit operations are performed after the reaction module and in equipment of similar design and capacity.

On the other hand, there is evidence that some process differences exist between the benzene and the n-butane process. The n-butane process is said to require a longer reactor residence time and, therefore, bigger reactors.[22] This means that a higher capital investment is required for the n-butane process to achieve the same production output, although there are few data upon which to estimate the increase in capital cost. One source estimates the capital costs for an n-butane plant could be 10 to 20 percent higher than for a comparable, benzene-based plant.[23] Also, production capacity would decrease for an existing benzene-based plant, if it were changed to use n-butane as a feedstock, unless the reactors were replaced with larger equipment. An additional problem is that the carbon steel reactors used in the oxidation of benzene may not withstand the higher temperatures of n-butane oxidation.[24] Furthermore, the composition of the stream from the reactors is likely to differ in the n-butane process, even though many of the same compounds will be present.[22] For this reason, variations in the design and operation of the product recovery and refining steps would be expected for a new plant based on n-butane.[19] The impact of this variation on the efficiency of a converted facility is not well known but could be a significant problem. According to Chemical Week:

> "One producer notes that the advisability of building new n-butane-based capacity versus converting old benzene-based capacity to n-butane is 'one of these questions not yet answered'."[21]

A switch in feeds for an existing facility would require a major investment in new catalysts;[8,21] for plants with an activated carbon adsorption unit attached to the product recovery absorber, additional modification costs would be incurred.[19] Moreover, the overall yield in converted plants would probably not be as high as in new plants designed specifically for n-butane; this lowered efficiency might be serious enough to warrant other major design changes.[21,22]

1.2.3 Byproduct of Phthalic Anhydride Production

Phthalic anhydride is manufactured from naphthalene and orthoxylene. Maleic anhydride is recovered as a byproduct from the plant effluent.[1] The emissions associated with maleic anhydride recovery are believed to be insignificant and are not being investigated at this time. There are no benzene emissions from this recovery process. In addition, no growth in the production of maleic anhydride by this process is expected.

1.2.4 Foreign Process

The only significant foreign process that differs from processes used in the United States is a process using feedstocks of 65 to 80 percent butenes with the remainder mostly n-butanes or isobutene (mixed C_4 oxidation). The general process description is similar to that shown in Figure 1-2 for benzene oxidation.[8] The exhaust from the main process vent contains unreacted n-butane, butene, carbon monoxide, and various secondary products. Except for the possible absence of benzene, the VOC emissions should be about the same as for the benzene oxidation process.[11]

The most significant process variation--the use of a fluidized catalyst bed rather than a fixed bed--was developed for the mixed C_4 process. This variation provides good temperature control within the bed, allowing optimum ratios of feed to air. In contrast, optimum feed-to-air ratios cannot be used with fixed bed systems because temperature cannot be precisely controlled; therefore, excess air becomes necessary to stay below the explosive range. The reduction of excess air in the fluidized bed feed will reduce emissions from the product recovery absorber. However, this advantage is offset by the lower product yields obtained with the fluidized bed than with the fixed bed.[4]

1.3 SUMMARY

This chapter has described the processes and associated emissions for four routes to the production of maleic anhydride. The benzene oxidation process is emphasized because it is the only one known to emit benzene. Two process emission points have been discussed: the product recovery absorber vent, and the refining vacuum system vent. Only the process emissions are discussed; emissions of benzene from fugitive and storage sources are to be addressed in another document that discusses these emissions from other chemical industries and petroleum refineries.

The n-butane oxidation process is discussed to the extent that available data permitted; it may replace benzene oxidation in the future.

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2. EMISSION CONTROL TECHNIQUES

2.1 ALTERNATIVE EMISSION CONTROL TECHNIQUES

This chapter addresses the control techniques that can be applied and the emission reductions that can be achieved for each benzene source identified in Chapter 1.

The control techniques that can be applied to reduce benzene emissions from maleic anhydride manufacturing facilities fall into two classes:

- Add-on control devices such as adsorbers or incinerators, and
- Use of another feedstock to eliminate benzene from the process.

For add-on control devices, it is assumed that the product recovery absorber vent and the refining vacuum system vent can be controlled by the same device. Only piping additions are required for controlling the refining vacuum system along with the product recovery absorber; no added utilities, manpower, or other operating costs are involved. Emissions from the refining vacuum system vent are included in all control system calculations. The control devices described here are applicable to the waste gas streams defined in Chapter 1. Furthermore, add-on control devices are expected to control the process vents during process upset conditions and startup (as discussed in Chapter 1) with no increase in emissions.

Because of the low concentrations of benzene and other volatile organic compounds (VOC) in the emissions from the main process vent (see Table 1-4), the use of flares is not generally believed to be an applicable control technique.[1] The high-volume, low-benzene concentration waste stream would require large amounts of auxiliary fuel without the opportunity for heat recovery.[2]

2.1.1 Carbon Adsorption

Carbon adsorption systems have often been applied in other industries to control waste gas streams carrying VOC for similar applications; e.g., with inlet organic concentrations ranging from 1 ppm to 40 volume percent and at air flow rates of 0.005 m^3/sec (11 cfm) to over 100 m^3/sec (212,000 cfm).[3] General discussions of the theory and application of carbon adsorption can be found elsewhere.[4,5,6]

Adsorption onto activated carbon can be used to recover benzene from gas streams from the product recovery absorber and refining vacuum system vent. Two plants currently use a carbon adsorber system to control the effluent from the product recovery absorber.[7] To use carbon adsorption in this application, the exhaust gas stream may be scrubbed with a caustic solution to remove organic acids and water-soluble organics. Benzene is probably the only VOC remaining in appreciable quantity after scrubbing. The waste stream is conditioned by reducing the relative humidity, which improves loading capacity.

Various levels of control can be achieved with carbon adsorption, depending on the design and operation of the adsorber system. In general, 99 percent or greater reduction of hydrocarbon emissions, including benzene, can be obtained.[7] Factors influencing the efficiency of carbon adsorption systems for benzene control in maleic anhydride plants include:

- Relative humidity of the incoming waste gas stream;
- Presence of other organic compounds that may interfere with benzene adsorption (possibly forming polymeric materials on the carbon beds) thereby decreasing capacity;
- Temperature of the beds and the gas during adsorption;
- Efficiency of the steam regeneration; and
- Dryness of the bed when put back on line.[8,9]

Both a caustic scrubber and a heater should be included in a carbon adsorption system, the caustic scrubber to remove most of the other organics in the stream and the preheater to lower the relative humidity of the water-saturated stream. Generally, a system of two or more carbon beds is used. The gas stream containing benzene passes through one or more beds in parallel, and the benzene is removed from the gas stream. At the same time, another bed is regenerated with low-pressure steam. The steam and desorbed benzene are condensed and decanted, after which the benzene returns to the process. The aqueous layer can be combined with the other liquid waste from the plant or recycled to the process. After regeneration, the carbon bed, which is hot and saturated with water, is usually cooled and dried, often by blowing ambient air through the bed. Bed size, number, and cycle times can be varied to achieve the desired removal efficiency. Finally, because the system may be exposed to corrosive compounds, stainless steel vessels are recommended. Specific systems that can achieve various removal efficiencies are discussed in Section 2.2.1. A carbon adsorption system, however, will not remove carbon monoxide, which is also present in the waste gas from the product recovery absorber.

2.1.2 Thermal Incineration

Thermal incineration, also called direct-fire incineration, can be used to control emissions from maleic anhydride manufacture. Three plants in the United States use a thermal incinerator on the product recovery absorber vent; one uses n-butane, while the other two use benzene as the feedstock.[7,10] General information on thermal incineration can be found elsewhere.[4,8] Because of the cost of the fuel required to operate a thermal incinerator, heat recovery is generally used. The recovered heat can be used either to preheat the feed to the incinerator or to generate steam. Some of the factors that influence the efficiency of incineration are temperature, degree of mixing, and residence time in the combustion chamber.

For maleic anhydride plants, a knockout demister tank is required to prevent liquid droplets from reaching the burner area.[7] Supplemental fuel is required to maintain necessary combustion temperatures, and supplemental combustion air may also be required if the incoming gas stream is not preheated. Because the gas stream to the incinerator contains corrosive materials, the equipment ahead of the combustion chamber must be made of stainless steel. Specific incineration systems and their removal efficiencies are described in Section 2.2.2.

The waste gas to be incinerated is approximately 2.5 mole percent fuel, which is below the combustible level (see Table 1-4). Thus, autoignition can only occur if the lower flammability limit of the mixture is 2.5 mole percent or less. Based on LeChatelier's Rule and the modified Burgess-Wheeler Law, the lower flammability levels of the waste gas mixture are 6.4 mole percent at 487° C (908° F) and 3.2 mole percent at 871° C (1,600° F). Consequently, autoignition of the waste gas does not occur even at the design temperature of the incinerator.[11]

Concern has been expressed that benzene emissions may leak through recuperative heat exchangers and incinerators. Except for rotary exchangers, heat exchangers generally do not leak. However, emissions are routinely prevented by maintaining the clean gas at a higher pressure than the side contaminated with benzene emissions. This higher pressure causes leakage into the feed stream to the incinerator rather than into the atmosphere. Shell and tube exchangers are not expected to leak. If routine maintenance is inadequate or operation is improper, leakage can occur because of ruptured tubes or warped flanges. However, this leakage is not caused by intrinsic equipment limitations.[12]

2.1.3 Catalytic Incineration

Catalytic incineration can be used as an alternative to thermal incineration of a waste gas stream. The catalyst allows oxidation to occur more rapidly at a lower temperature, thereby decreasing or eliminating the need for supplemental fuel consumption. A catalytic incinerator is currently used by one maleic anhydride producer to meet local hydrocarbon regulations. The unit removes at least 85 percent of the total hydrocarbon content and 95 percent of the CO content of the product recovery absorber vent gas without additional fuel.[13]

Theoretically, a catalytic incinerator could be designed to have a removal efficiency of 90 to 95 percent.[8] In practice, however, no such system is in operation on a maleic anhydride production facility because cost may be prohibitive. Achieving high benzene removal efficiencies typically requires large catalyst volumes or temperatures close to those of thermal incineration, so a catalytic converter is believed to be uneconomical.[8]

In addition, depending on the nature of the catalyst used, fouling of the catalyst may reduce its life. This may occur in maleic anhydride plants because some of the components of the waste gas stream may polymerize.[8] One catalyst producer believes this problem can be overcome by using a monolithic support system with platinum catalysts and periodic maintenance.[14] Precious metal catalysts are not deactivated by water. The deposition of high-boiling organics on the catalyst's surface merely masks that surface. Periodic maintenance is often all that is required to alleviate this problem.[14]

2.1.4 n-Butane Process Conversion

The n-butane oxidation process has the potential for zero benzene emissions, permitting conversion from benzene feed to n-butane feed to be considered a control technique for reducing benzene emissions from the manufacture of maleic anhydride. The process, emissions, and factors influencing conversion from benzene to n-butane are discussed in Chapter 1. One company using n-butane has detected no benzene emissions from its reactors at the lower detection limits of 1 ppmv.[15] In addition, no benzene has been detected at the 0.5-ppm level in the final product from the n-butane process.[16] Problems associated with conversion to n-butane were discussed in Section 1.2.2.

2.2 PERFORMANCE OF EMISSION CONTROL TECHNIQUES

2.2.1 Carbon Adsorption

Based on engineering experience with similar applications for the control of VOC, a carbon adsorption system can be designed and operated at a sustained benzene removal efficiency of 99 percent.[7] This efficiency reflects optimal control of temperature, pressure, humidity, and the level of other organics. Carbon adsorption systems operating at different levels of control are described in this section. The first system, at 99 percent control of benzene, is based on engineering design calculations and previous engineering experience. The other systems described here are based on the experience of existing maleic anhydride facilities, one system reported by the company to achieve an average of only 85 percent control of benzene, while the efficiency of the other has not been determined.[17]

As discussed in Section 2.1.1, several key factors influence the efficiency of the adsorption system, and optimum control of these factors is responsible for 99 percent reduction of benzene emissions. The carbon adsorption system uses a caustic scrubber. A heater is used before the carbon beds to decrease the relative humidity.

The size of the activated carbon adsorbers is determined by the superficial velocity through the bed and the carbon requirements, as determined by the flow of adsorbable species and the carbon loading for each species. The superficial velocity is important because it affects the pressure drop through the bed. The usually acceptable range for superficial velocity is 25 to 50 cm/sec (10 to 20 in/sec), which gives a pressure drop of 25 to 65 cm H_2O/m (3 to 8 in H_2O/ft) bed. The air flow rate to the carbon adsorber system is approximately 21 m^3/sec (4.5×10^4 cfm). Assuming a superficial velocity of 38 cm/sec (15 in/sec), the required cross-sectional area is about 55 m^2 (590 ft^2). Because adsorbers this large are usually horizontal tanks, two tanks 3 m (10 ft) diameter × 9 m (30 ft) long operated in parallel would give the required cross-sectional area.

To allow enough time for regeneration, additional adsorbers are necessary. To ensure that 99 percent removal could be achieved for the system, which is designed for comparison and costing purposes, two additional adsorbers were chosen instead of one. Also, experience from one adsorber system in this application suggests there is not sufficient time to complete the regeneration cycle (including the cooling and drying cycle) and thereby prevent breakthrough when only three beds are used.[17]

The amount of carbon in the adsorber is determined by the desired cycle time, flow rate of adsorbable species, and carbon loading. Although shorter cycle times have been used, a 2-hour loading cycle was chosen initially to allow a 1-hour steaming cycle and another hour to cool and dry the carbon in preparation for adsorption service. The amount of carbon per adsorber can be calculated from emission rate, loading, and bulk density:

$$\frac{\text{kg (lb) carbon}}{\text{bed cycle}} = \frac{190\ \text{kg (418 lb) Bz}}{\text{hr}} \times \frac{2\ \text{hr}}{\text{cycle}}$$

$$\times \frac{1\ \text{kg (2.2 lb) carbon}}{0.06\ \text{kg (0.13 lb) Bz}} \times \frac{1}{2\ \text{beds}}$$

$$\frac{\text{kg (lb) carbon}}{\text{bed cycle}} = 3{,}200\ \text{kg (7,000 lb)}$$

$$\text{Volume of carbon} = 3{,}200\ \text{kg (7,000 lb)} \times \frac{\text{m}^3\ (35.3\ \text{ft}^3)}{448\ \text{kg (985.6 lb)}}$$

$$= 7.1\ \text{m}^3\ (250\ \text{ft}^3)\quad .$$

The depth of the carbon required to give a 2-hour loading cycle is 25 cm (10 in) in the adsorbers selected. Thus, in this case, the size of the adsorber is dictated more by the total flow rate than by the carbon requirements. Because the incremental cost of putting more carbon in the adsorbers is small, a carbon depth of 61 cm was used. Each adsorber would now hold 17 m^3 (600 ft^3) or 7,600 kg (16,800 lb) of carbon. Using the same carbon loading and benzene flow as before now gives an adsorption cycle of 4.8 hours, which should allow enough adsorption capacity to prevent premature breakthrough while the companion bed is being regenerated.

A loading of 0.06 kg benzene (Bz) per kilogram (0.06 lb Bz/lb) of carbon has been selected based on capacity data reported by Hoyt Manufacturing (a carbon adsorber vendor).[9] These capacity data are also similar to Reichhold's experience with a benzene adsorber in this application.[18] The selected loading is conservative because other data in the literature show benzene loadings as high as 0.25 kg Bz/kg (0.25 lb Bz/lb) of carbon on successive cycles for full-scale systems adsorbing benzene from town gas.[19] If higher loadings are achieved, the loading cycle would be extended, allowing still more time for regeneration, cooling, drying, and lower operating costs.

There is some indication that the relative humidity of the vent stream can affect the loading capacity. Hydroscience has observed a 75- to 80-percent drop in loading for some materials at a relative humidity of 80 percent as compared to 20 percent relative humidity.[9,20] Similar data for activated carbon are reported by Rohm and Haas in their technical bulletin on Ambersorb Carbonaceous Resins.[21] It is important to note, however, that changes were noted only in the loading capacity and not in the ability to achieve a baseline outlet concentration of less than 5 ppmv.[9]

To regenerate the carbon beds 20 kg steam/kg Bz (20 lb steam/lb Bz) are required. This amount should be more than adequate to complete the desorption because steam requirements as low as 3 kg steam/kg Bz (3 lb steam/lb Bz) toluene have been quoted in the literature.[22] Some flexibility is available when the duration of the steaming cycle is determined; steaming can easily be completed in 1 hour or less if the condensing capacity exists. For a new installation, it is best to size the condenser to handle the steam flow equivalent for at least a 1-hour steam cycle as an additional safety factor. The average steam flow (for calculating annualized costs) of 1.1 kg/sec (2.4 lb/sec) is determined from the average benzene flow and the steam-to-benzene ratio.

An important part of the total operation is the cooling and drying sequence. If the bed is not properly cooled and dried, both poor efficiency and lower adsorption capacities will result. Note that this result contrasts with the previously discussed effect of relative humidity of the inlet gas on loading capacity and efficiency. The full-scale experience with benzene adsorbers at one facility where cooling and drying are not practiced shows that there was an initial spike of benzene in the outlet immediately after the hot bed was put back on line.[17] In addition to this initial spike, the instantaneous removal was rarely better than 97 percent. These data are consistent with the observations that poor efficiency and reduced loading capacity occur when the adsorber is not properly cooled and dried following a steam regeneration. In order to achieve greater than 99 percent removal, the bed must be cooled and dried. In addition, the exhaust from the cooling and drying cycle must be recycled to a carbon bed that is in the adsorption mode. This means that the adsorbers must be sized so there is capacity to handle the additional benzene and air flow as well as time to complete the cooling and drying cycle. The data from one facility show that the cooling and drying period lasted for about the first 15 minutes of waste gas incineration and that about 1 to 3 percent of the benzene in the feed left the system during this period.[17] If the worst case of 3 percent is used, the calculated adsorber duty would be about 197 kg/hr (433 lb/hr) instead of 190 kg/hr (418 lb/hr). The calculated time to breakthrough (at a loading of 0.06 kg Bz/kg carbon [0.06 kg Bz/kg]) would only be reduced from 4.8 hours to 4.65 hours.

The effect of total flow rate on efficiency and pressure drop through the on-line adsorber is more critical than the recycle of some adsorbed benzene since outside air must be used for drying. The vent itself is not suitable for drying (even though the relative humidity of the vent downstream of the scrubber is reduced from 100 percent relative humidity to about 50 percent relative humidity by the preheater).[9] For design purposes, a cooling and drying cycle of 1 hour at a flow rate equivalent to one-fourth to one-half of the flow rate during the adsorbing cycle was chosen. This would raise the superficial velocity for the adsorbing carbon bed from 38 cm/sec (15 in/sec) to 49 to 57 cm/sec (19 to 22 in/sec) during the cooling and drying cycle but would not create an excessive pressure drop. Because two beds could be cooling at once, the blower has been sized to deliver 21 m^3/sec (4.5×10^4 cfm). A capital savings could be achieved if a blower were sized at 11 m^3/sec (2.3×10^4 cfm). However, the additional capital cost of the larger blower is considered justified to make the operation of the adsorption system more flexible. Pertinent details of the system designed to achieve 99 percent control of benzene are in Table 2-1, and costs for such systems are presented for each plant in Chapter 5.

As indicated previously, two plants currently use carbon adsorption systems to control benzene from product recovery absorber vents. One plant

TABLE 2-1. TECHNICAL DATA--CARBON ADSORPTION SYSTEM
(99 Percent Control)

Number of beds	4
Weight of carbon per bed	7,620 kg (16,800 lb)
Loading cycle time	2 hr
Number of stacks	2
Benzene emission rate per stack	0.26 g/sec (5.8×10^{-4} lb/sec)
Air flow rate from system (per stack)	18.0 m^3/sec (3.8×10^4 cfm)

NOTE: References and basis explained in text.

TABLE 2-2. TECHNICAL DATA--CARBON ADSORPTION SYSTEM
(85 Percent Control)

Number of beds	3
Loading cycle time	2 hr
Regeneration cycle time	1 hr
Number of stacks	1
Air flow rate from system[a]	20.3 m^3/sec (4.3×10^4 cfm)

[a] At capacity.

SOURCE: Lawson, J. F. Trip Report for Visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977. Hydroscience, Inc. EPA Contract Number 68-02-2577.

for which data are available reports that it has achieved a sustained benzene removal efficiency of only 85 percent.[17] Results of tests conducted by the U.S. Environmental Protection Agency (EPA) showed a mean benzene removal efficiency of 93 percent (with a 90-percent confidence interval of 89 to 97 percent).[23] However, the plant was operating at only 40 percent of full capacity when tested. Pertinent technical data on that system are summarized in Table 2-2. This system does not use an organic-free gas stream to cool and dry the beds after regeneration with steam. Immediately after regeneration, the waste gas stream containing benzene is directed to the hot bed. Consequently, until the bed cools benzene removal efficiency is low, which partially accounts for the low overall benzene removal efficiency.

The system at the second plant was designed for a benzene removal efficiency of 98 percent, although no reliable data are currently available upon which to estimate its performance. The system can be expected to show better performance than the one discussed previously because the beds are cooled and dried with air after steam desorption.[24] Pertinent technical data on this system are in Table 2-3.

2.2.2 Thermal Incineration

Based on engineering experience with similar applications for the control of VOC's, it is expected that a thermal incinerator can be designed and operated at a sustained benzene removal efficiency of 99 percent.[7] Limited information is available on direct-flame afterburners used on maleic anhydride production facilities, but there are several cases in which streams similar to the product recovery absorber and refining vacuum system vent gas have been controlled at high efficiencies. In one case, data reported for toluene indicate a destruction efficiency of 99.9 percent at 766° C (1,411° F) and a residence time of 0.21 second for toluene.[25] Another facility incinerates a toluene-xylene vapor at 760° C (1,400° F) and reportedly achieves a destruction efficiency of 99.1 percent.[26] A third installation reportedly expects a destruction efficiency of greater than 99 percent at 760° C (1,400° F) for an organic stream considered as toluene.[26] Finally, a review of several case studies indicates that combustion efficiencies of less than 95 percent were achieved, except in one case, at temperatures of 730° C (1,346° F) or lower. Conversely, efficien-

TABLE 2-3. TECHNICAL DATA--CARBON ADSORPTION SYSTEM
(98 Percent Control at Design)

Number of beds	2
Loading cycle time	6 hr
Regeneration cycle time	1.5 hr
Cooling and drying cycle time	0.75 hr
Number of stacks	2
Waste gas flow rate[a]	11.3 m^3/sec (2.4 × 10^4 cfm)
Cooling air flow rate	11.3 m^3/sec (2.4 × 10^4 cfm)
Weight of carbon per bed	15,900 kg (35,000 lb)

[a] At capacity.

SOURCE: Weber, Robert. Trip Report for Visit to Reichhold Chemicals, Inc., Elizabeth, N.J., July 15, 1978. U.S. Environmental Protection Agency.

cies of 99 percent plus were achieved at temperatures of 760° C (1,400° F) or greater.[27]

Research data on a fume incinerator also indicate high removal efficiencies for similar streams. With a toluene-contaminated gas, it was found that approximately 99 percent destruction efficiency was achieved at 760° C (1,400° F) and a residence time of 0.33 second. As the temperature increased to 816° C (1,501° F), the efficiency increased to 99.5 percent.[27]

Recent laboratory studies on thermal incineration of benzene also show high destruction efficiencies.[28] With a detection limit of 2 ppmv, no benzene was found at temperatures above 790° C (1,454° F) with residence times as low as 0.08 second. Based on that research, a thermal oxidizer with a residence time of 0.5 second would require a temperature of 750° C (1,382° F) for 99.9 percent destruction of benzene.

One plant controls the emissions from the product recovery absorber by routing the waste gas stream to a waste heat boiler. In tests conducted in 1977, the average benzene removal efficiency was calculated to be 99 percent with benzene outlet concentrations ranging from 6.0 ppmv to 9.0 ppmv.[29] Although the operating temperature is not reported, it is probably near 2,000° F (1,090° C). In later tests in 1978, the benzene removal efficiency was estimated to range between 95.8 percent and 98.5 percent, with benzene outlet concentrations ranging from 8.5 ppmv to 13 ppmv.[30] Further, it was stated that ". . . the boiler now operates somewhat below 2,000° F."[30] During this set of tests, the inlet benzene concentrations were not measured. A waste heat boiler, however, is a viable option only if the facility can economically use the additional steam generated.

A temperature of 870° C (1,600° F) is specified to ensure complete combustion of the waste gas. While it is possible that greater than 99 percent VOC removal can be obtained at lower temperatures, it cannot be predicted dependably. The conditions of Table 2-4 are consistent with various air pollution engineering manuals.[5,31] While the manuals do not provide data on combustion temperatures above 800° C (1,472° F), extrapolation of the data presented combined with the similar incineration experience described above supports the projection of greater than 99 percent removal at 870° C (1,600° F). The costs for systems of this type are presented in Chapter 5 for each maleic anhydride plant.

Based on both the data from the maleic anhydride facility with the boiler and the data discussed above on similar incinerator systems, removal efficiency of 99 percent for benzene at 870° C (1,600° F) is anticipated in a well-designed and efficiently operated incinerator. Pertinent technical data are summarized in Table 2-4.

A second plant uses a thermal incinerator on this stream operating at 760° C (1,400° F) with a residence time of 0.7 second.[32] Total hydrocarbon destruction efficiency has been reported to be at least 93 percent, although the company has no data on benzene removal efficiency. Tests conducted by EPA showed a mean benzene removal efficiency of 98.6 percent.[33] Although the plant was operating at about 70 percent of capacity when sampling was conducted, the plant personnel did not think the lower production rate would seriously affect the validity of the results when applied at full capacity. Pertinent technical data are summarized in Table 2-5. This incinerator is also used to generate steam, with a steam production rate of about 7 kg/sec (15.4 lb/ sec) during the testing.[34]

Generally, an incinerator can operate at least 95 percent of the time with proper attention to maintenance and operating parameters.[12] Even if repairs are required, they can often be made quickly or postponed until the next plant shutdown.

As discussed previously, the process waste gases are destroyed in an onsite boiler in one facility; however, other maleic anhydride plants have high gas volumes that exceed the total air-firing needs of those plants' steam boilers. (As discussed earlier in this chapter and Chapter 1, these high values are required to maintain the percent by volume of benzene in the feedstock to 1.5 or less. This leads to large volumes of waste gases with low benzene concentrations.) Consequently, most plants cannot handle the waste gases in onsite waste heat boilers because the existing equipment is not designed for such high-volume, low-concentration gases.[2]

In addition to controlling VOC emissions, thermal incineration will reduce CO concentrations. At a temperature of 609° C (1,128° F), thermal incineration converts 90 percent of the CO to CO_2.[35] For example, at the higher temperatures of a plant with 99 percent benzene removal, 870° C (1,600° F), more than 99 percent of the CO is expected to be oxidized.

TABLE 2-4. DESIGN CRITERIA--THERMAL INCINERATION
(99 Percent Control)

Residence time	0.5 sec
Temperature	870° C (1,600° F)
Natural gas (supplemental fuel)	0.435 m^3/sec (922 cfm)
Supplemental combustion air	1.24 kg/sec (2.7 lb/sec)

NOTE: References and basis explained in text.

TABLE 2-5. TECHNICAL DATA--THERMAL INCINERATION
(97 Percent Control)

Residence time	0.7 sec
Temperature	760° C (1,400° F)
Natural gas (supplemental fuel)	0.5 m^3/sec (1,066 cfm)
Supplemental combustion air[a]	6.5 kg/sec (14.3 lb/sec)

[a] DENKA's incinerator is also used to produce steam for various processes. The air used beyond theoretical combustion requirements may be needed to produce enough steam for these processes.

SOURCE: Weber, Robert. Trip Report for Visit to DENKA Chemical Corporation, Houston, Texas, April 20, 1978. Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency.

2.2.3 Catalytic Incineration

One maleic anhydride producer reports VOC removal of at least 85 percent and carbon monoxide removal of greater than 95 percent, with a catalytic incinerator on the absorber offgas.[13] In tests run on a pilot catalytic incinerator with a nominal capacity of 1.4 m^3/sec (3,000 scfm) of maleic anhydride exhaust gases, benzene control was about 98.7 percent, and carbon monoxide control was about 99.3 percent.[36] These tests were based on approximately 300 hours of operating time.

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3. REGULATORY OPTIONS

This chapter describes three regulatory options for two emission points in maleic anhydride production facilities: the product recovery absorber vent, and the refining vacuum system vent. The same control device can be used to reduce emissions from both sources.

For existing maleic anhydride production facilities, two control levels are considered viable alternatives for a standard based on best available technology (considering environmental, energy, and economic impacts). These options, and examples of applicable control techniques, are:

- Benzene control efficiency of 97 percent, which can be achieved by carbon adsorption or thermal incineration; and
- Benzene control efficiency of 99 percent, which can be achieved by either thermal incineration or carbon adsorption.

The control devices; i.e., incinerators and carbon adsorbers, can be designed and operated at various levels of control efficiency. The benzene control efficiency of 97 percent was chosen to represent universally applicable control already demonstrated on a maleic anhydride facility. Although tests conducted at this facility indicated a 98-percent removal efficiency, they were conducted at a production rate of only 70 percent of capacity. Therefore, because it is uncertain whether the same control efficiency would have been obtained if the plant were operating at 100 percent of capacity, 97 percent benzene removal was selected as the basis for one regulatory option. The 99-percent removal efficiency is based on technology transfer, as described in Chapter 2.

As discussed in Chapter 1, uncertainty continues regarding the viability of converting existing benzene-based plants to facilities using n-butane as the feedstock. Little information is available on what would be required to convert each existing benzene-based plant to an n-butane-based plant, or what the consequences of such a conversion would be. Based on the limited information available, it appears that considerable effort continues to be directed towards developing n-butane technology, particularly the catalyst. Only the existing Amoco plant was originally designed to use n-butane as a feedstock. Problems associated with converting existing plants to n-butane include:

- Potentially significant reductions in maleic anhydride production when current n-butane catalyst technology is combined with equipment designed for benzene as the feedstock, and
- Unsatisfactory operation resulting from equipment changes needed in the refining system.

Because of uncertainties concerning the technical feasibility of converting each existing source to n-butane and the impacts of such conversion, this approach is not considered a viable regulatory option for existing sources based on BAT.

The use of n-butane as a feedstock, however, is considered a viable regulatory option for new sources. Because the industry was operating at only 56 percent of capacity in 1978, few new sources are expected to be built until the mid-1980's. This allows time for continued development of the n-butane process. Furthermore, a new plant could be designed to use n-butane and would therefore not encounter the potential problems associated with conversion. In fact, one company has recently announced the construction of a new 45,400-Mg/yr (50,000 tons/yr) maleic anhydride plant based on their proprietary n-butane technology, which is scheduled for completion in early 1983.

In summary, only two of the three options outlined above were considered viable regulatory options that could serve as the basis for a standard for existing sources based on BAT. These two regulatory options are designated: Option 1, 97 percent benzene control, and Option 2, 99 percent benzene control. For new sources, a regulatory option of no detectable benzene emissions (100 percent control) is considered a viable option as the basis for BAT. One hundred percent control was also considered as an option beyond BAT for existing sources. The environmental and energy impacts of these alternatives for existing and new sources are presented in Chapter 4 and in the economic impact section of Chapter 5.

4. ENVIRONMENTAL AND ENERGY IMPACT

The environmental and energy impacts of each regulatory option presented in Chapter 3 are discussed in this chapter. Both beneficial and adverse environmental impacts are assessed. The two regulatory options considered for existing plants are:

- A removal efficiency for benzene of 97 percent based on carbon adsorption or thermal incineration, and
- A removal efficiency for benzene of 99 percent based on carbon adsorption or thermal incineration.

Catalytic incineration is not assessed as a control technique at this time because control efficiencies in the 97-percent to 99-percent range have not been demonstrated on a commercial scale. However, if these efficiencies could be achieved, catalytic incineration would be a desirable control technique because it would:

- Operate with essentially no fuel consumption,
- Produce no secondary emissions,
- Provide for changing catalyst if required by further regulations,
- Oxidize CO to CO_2, and
- Probably have lower capital costs.[1]

The additional regulatory option for new plants of elimination of benzene by substituting n-butane, or similar material, as the feedstock is also assessed. Both primary and secondary impacts of these regulatory options are discussed in the next sections.

4.1 AIR POLLUTION IMPACT

This section addresses both the positive and negative effects on air pollution expected to result from the application of the regulatory options. The emission rates from a model plant as described in Chapter 1 are presented in Table 4-1 for the following cases:

TABLE 4-1. EMISSION RATES AND MAXIMUM ANNUAL AVERAGE BENZENE CONCENTRATIONS[a,b]

Model plant control alternatives	Emission rate in g/sec (lb/sec)[c]	Combined maximum ground level concentration in μg/m³ (ppmv)	Contribution of the process vents to the combined maximum concentration of all sources μg/m³ (ppmv)
A. Uncontrolled model plant	96.0 (2.12×10^{-1})	60.7 (1.9×10^{-2})	56.4 (1.8×10^{-2})
B. 99 percent control by incineration	0.96 (2.12×10^{-3})	20.7 (6.5×10^{-3})	1.5×10^{-2} (4.7×10^{-6})
C. 99 percent control by carbon adsorption	0.48 (1.06×10^{-3}) (Stack A) 0.48 (1.06×10^{-3}) (Stack B)	20.8 (6.5×10^{-3})	1.4×10^{-1} (4.4×10^{-5})
D. 97 percent control by carbon adsorption	2.88 (6.35×10^{-3})	20.7 (6.5×10^{-3})	5.8×10^{-2} (1.8×10^{-5})

[a] Storage, handling, and fugitive emissions are assumed to be uncontrolled in all cases.

[b] Based on a benzene conversion rate of 90 percent.

[c] Only refers to emission rate from the product recovery absorber and vacuum system vent or to the control device for these emission points.

SOURCE: H. E. Cramer Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions From a Maleic Anhydride Plant for Four Emission Control Options. Salt Lake City, Utah. (Prepared for Source-Receptor Analysis Branch, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2507. August 1978.

A. Uncontrolled model plant,

B. 99 percent control by thermal incineration,

C. 99 percent control by carbon adsorption, and

D. 97 percent control by carbon adsorption.

Table 4-1 also includes the estimated maximum ambient ground level concentrations of benzene in the vicinity of the model facilities, determined by atmospheric dispersion modeling.[2] This atmospheric dispersion model was originally run for 97 and 99.5 percent control levels. Thus, the data given for 99 percent control are revised from the original data in Reference 2 to reflect a change in control level from 99.5 to 99 percent. This effectively doubles both the amount of benzene emitted and the level of ambient benzene concentrations. Table 4-2 shows the mean annual average ground level benzene concentrations produced by the emissions from combined fugitive, storage, and process sources based on the atmospheric dispersion model.[2] Table 4-3 summarizes the mean annual benzene concentrations produced by emissions from storage, fugitive, and process sources; it is based on the atmospheric dispersion model[2] with the following assumptions:

- Storage tanks A and B are controlled to the 90-percent level.
- Fugitive sources are controlled to the 90-percent level.
- Emissions from the product recovery scrubber, the incinerator vent, Vents A and B of the adsorber system, and the adsorption system vent are assumed to remain the same with or without fugitive and storage controls.
- Percentage of total emissions is the percentage of the total emitted for that particular source, storage, fugitive, or process, and option combination.

This table shows the differences in ambient concentrations with and without fugitive and storage emissions control; fugitive and storage emissions control may be considered in the future.

The Industrial Source Complex (ISC) Dispersion Model was used to evaluate the air quality impact of benzene emissions from the model maleic anhydride plant. The program was used to calculate the maximum annual average ground level benzene concentration resulting from each model plant control alternative, for both the combined emissions from all sources of benzene in the plant as well as the concentration from the product recovery absorber and the refining vacuum system vent (both uncontrolled and controlled). This information is presented in Table 4-1. The program also

TABLE 4-2. MEAN ANNUAL AVERAGE BENZENE CONCENTRATIONS PRODUCED BY EMISSIONS FROM THE COMBINED SOURCES[a]

Model plant control alternative	Mean concentration ($\mu g/m^3$)									
	0.1 km	0.2 km	0.3 km	0.5 km	0.7 km	1.0 km	1.5 km	2.0 km	10.0 km	20.0 km
A	1.71×10^1	2.42×10^1	3.22×10^1	3.43×10^1	3.04×10^1	2.48×10^1	1.74×10^1	1.29×10^1	1.39	5.03×10^{-1}
B	1.28×10^1	6.36	3.95	2.09	1.35	8.58×10^{-1}	5.05×10^{-1}	3.46×10^{-1}	3.69×10^{-2}	1.11×10^{-2}
C	1.30×10^1	6.77	4.46	2.48	1.60	9.89×10^{-1}	5.64×10^{-1}	3.73×10^{-1}	3.13×10^{-2}	1.10×10^{-2}
D	1.29×10^1	6.77	4.63	2.85	1.98	1.37	8.48×10^{-1}	5.89×10^{-1}	5.63×10^{-2}	2.02×10^{-2}

[a] Concentration in ppmv = (0.000314) × concentration in $\mu g/m^3$ (at 25° C and at 1 atm).

SOURCE: H. E. Cramer Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions from a Maleic Anhydride Plant for Four Emission Control Options. Salt Lake City, Utah. (Prepared for Source-Receptor Analysis Branch, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) August 1978.

TABLE 4-3. MEAN ANNUAL BENZENE CONCENTRATIONS PRODUCED BY EMISSIONS FROM INDIVIDUAL SOURCES

	0.1 km				0.2 km			
	No fugitive or storage controls		Fugitive and storage controls^{a,b}		No fugitive or storage controls		Fugitive and storage controls^{a,b}	
Source/alternative	Concentration^d µg/m^3	% of total emissions^c	Concentration^d µg/m^3	% of total emissions^c	Concentration^d µg/m^3	% of total emissions^c	Concentration^d µg/m^3	% of total emissions^c
Storage tank A								
A	5.57	32.6%	5.57×10^{-1}	10.0%	3.09	12.8%	3.09×10^{-1}	1.7%
B	5.57	43.5%	5.57×10^{-1}	42.5%	3.09	48.6%	3.09×10^{-1}	41.2%
C	5.57	42.8%	5.57×10^{-1}	36.9%	3.09	45.6%	3.09×10^{-1}	26.0%
D	5.57	43.2%	5.57×10^{-1}	39.2%	3.09	45.6%	3.09×10^{-1}	26.9%
Storage tank B								
A	1.39	8.1%	1.39×10^{-1}	2.5%	5.50×10^{-1}	2.3%	5.50×10^{-2}	0.3%
B	1.39	10.9%	1.39×10^{-1}	10.6%	5.50×10^{-1}	8.7%	5.50×10^{-2}	7.4%
C	1.39	10.7%	1.39×10^{-1}	9.2%	5.50×10^{-1}	8.1%	5.50×10^{-2}	4.6%
D	1.39	10.8%	1.39×10^{-1}	9.8%	5.50×10^{-1}	8.1%	5.50×10^{-2}	4.8%
Fugitive sources								
A	5.75	33.6%	5.75×10^{-1}	10.3%	2.60	10.7%	2.60×10^{-1}	1.4%
B	5.75	44.9%	5.75×10^{-1}	43.9%	2.60	40.9%	2.60×10^{-1}	35.0%
C	5.75	44.2%	5.75×10^{-1}	38.1%	2.60	38.4%	2.60×10^{-1}	21.8%
D	5.75	44.6%	5.75×10^{-1}	40.5%	2.60	38.4%	2.60×10^{-1}	22.6%
Product recovery scrubber								
A	4.29	25.1%	4.29	77.2%	1.79×10^{1}	74.0%	1.79×10^{1}	96.8%
Incinerator vent								
B	3.89×10^{-2}	0.3%	3.89×10^{-2}	2.97%	1.18×10^{-1}	1.86%	1.18×10^{-1}	15.9%
Adsorber system vent A								
C	1.19×10^{-1}	0.93%	1.19×10^{-1}	7.9%	2.71×10^{-1}	4.0%	2.71×10^{-1}	22.8%
Adsorber system vent B								
C	1.14×10^{-1}	0.89%	1.14×10^{-1}	7.6%	2.59×10^{-1}	3.8%	2.59×10^{-1}	21.8%
Adsorption system vent								
D	1.46×10^{-1}	1.13%	1.46×10^{-1}	10.3%	5.22×10^{-1}	7.7%	5.22×10^{-1}	45.4%
TOTALS								
A	1.71×10^{1}		5.56		2.42×10^{1}		1.85×10^{1}	
B	1.28×10^{1}		1.31		6.36		7.42×10^{-1}	
C	1.30×10^{1}		1.51		6.77		1.19	
D	1.29×10^{1}		1.42		6.77		1.15	

Notes are found at the end of Table 4-3.

(continued)

TABLE 4-3. (continued)

Source/alternative	0.3 km				0.5 km			
	No fugitive or storage controls		Fugitive and storage controls[a,b]		No fugitive or storage controls		Fugitive and storage controls[a,b]	
	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]
Storage tank A								
A	1.96	6.1%	1.96×10^{-1}	0.7%	9.90×10^{-1}	2.9%	9.90×10^{-2}	0.3%
B	1.96	49.6%	1.96×10^{-1}	33.9%	9.90×10^{-1}	47.4%	9.90×10^{-2}	21.7%
C	1.96	44.0%	1.96×10^{-1}	18.0%	9.90×10^{-1}	40.0%	9.90×10^{-2}	11.8%
D	1.96	42.3%	1.96×10^{-1}	15.6%	9.90×10^{-1}	34.7%	9.90×10^{-2}	8.2%
Storage tank B								
A	3.18×10^{-1}	0.1%	3.18×10^{-2}	0.1%	1.47×10^{-1}	0.4%	1.47×10^{-2}	0.04%
B	3.18×10^{-1}	8.1%	3.18×10^{-2}	5.5%	1.47×10^{-1}	7.0%	1.47×10^{-2}	3.2%
C	3.18×10^{-1}	7.1%	3.18×10^{-2}	2.9%	1.47×10^{-1}	5.9%	1.47×10^{-2}	1.8%
D	3.18×10^{-1}	6.9%	3.18×10^{-2}	2.5%	1.47×10^{-1}	5.2%	1.47×10^{-2}	1.2%
Fugitive sources								
A	1.47	4.6%	1.47×10^{-1}	0.5%	6.77×10^{-1}	2.0%	6.77×10^{-2}	0.2%
B	1.47	37.2%	1.47×10^{-1}	25.4%	6.77×10^{-1}	32.4%	6.77×10^{-2}	14.9%
C	1.47	33.0%	1.47×10^{-1}	13.5%	6.77×10^{-1}	27.3%	6.77×10^{-2}	8.1%
D	1.47	31.8%	1.47×10^{-1}	11.7%	6.77×10^{-1}	23.8%	6.77×10^{-2}	5.6%
Product recovery scrubber								
A only	2.85×10^{1}	88.5%	2.85×10^{1}	98.6%	3.25×10^{1}	94.8%	3.25×10^{1}	99.4%
Incinerator vent								
B only	2.03×10^{-1}	5.14%	2.03×10^{-1}	35.1%	2.7×10^{-1}	12.9%	2.7×10^{-1}	59.2%
Adsorber system vent A								
C only	3.6×10^{-1}	8.07%	3.60×10^{-1}	33.0%	3.32×10^{-1}	13.4%	3.32×10^{-1}	39.5%
Adsorber system vent B								
C only	3.55×10^{-1}	7.96%	3.55×10^{-1}	32.6%	3.31×10^{-1}	13.3%	3.31×10^{-1}	39.4%
Adsorption system vent								
D only	8.82×10^{-1}	19.1%	8.82×10^{-1}	70.2%	1.03×10^{-1}	36.2%	1.03×10^{-1}	85.4%
TOTALS								
A	3.22×10^{1}		2.89×10^{1}		3.43×10^{1}		3.27×10^{1}	
B	3.95		5.78×10^{-1}		2.09		4.56×10^{-1}	
C	4.46		1.09		2.48		8.41×10^{-1}	
D	4.63		1.26		2.85		1.21	

(continued)

TABLE 4-3. (continued)

Source/alternative	0.7 km				1.0 km			
	No fugitive or storage controls		Fugitive and storage controls[a,b]		No fugitive or storage controls		Fugitive and storage controls[a,b]	
	Concentration[d] µg/m^3	% of total emissions[c]	Concentration[d] µg/m^3	% of total emissions[c]	Concentration[d] µg/m^3	% of total emissions[c]	Concentration[d] µg/m^3	% of total emissions[c]
Storage tank A								
A	6.0×10^{-1}	2.0%	6.0×10^{-2}	0.2%	3.51×10^{-1}	1.4%	3.51×10^{-2}	0.14%
B	6.0×10^{-1}	44.4%	6.0×10^{-2}	15.4%	3.51×10^{-1}	40.9%	3.51×10^{-2}	11.6%
C	6.0×10^{-1}	37.5%	6.0×10^{-2}	9.5%	3.51×10^{-1}	35.9%	3.51×10^{-2}	8.1%
D	6.0×10^{-1}	30.3%	6.0×10^{-2}	5.9%	3.51×10^{-1}	25.6%	3.51×10^{-2}	4.3%
Storage tank B								
A	8.53×10^{-2}	0.28%	8.53×10^{-3}	0.03%	4.87×10^{-2}	0.2%	4.87×10^{-3}	0.02%
B	8.53×10^{-2}	6.3%	8.53×10^{-3}	2.2%	4.87×10^{-2}	5.7%	4.87×10^{-3}	1.6%
C	8.53×10^{-2}	5.3%	8.53×10^{-3}	1.4%	4.87×10^{-2}	4.9%	4.87×10^{-3}	1.1%
D	8.53×10^{-2}	4.3%	8.53×10^{-3}	0.8%	4.87×10^{-2}	3.6%	4.87×10^{-3}	0.6%
Fugitive sources								
A	3.91×10^{-1}	1.3%	3.91×10^{-2}	0.13%	2.16×10^{-1}	0.87%	2.16×10^{-2}	0.09%
B	3.91×10^{-1}	29.0%	3.91×10^{-2}	10.0%	2.16×10^{-1}	25.2%	2.16×10^{-2}	7.1%
C	3.91×10^{-1}	24.4%	3.91×10^{-2}	6.2%	2.16×10^{-1}	21.8%	2.16×10^{-2}	5.0%
D	3.91×10^{-1}	19.8%	3.91×10^{-2}	3.8%	2.16×10^{-1}	15.8%	2.16×10^{-2}	2.7%
Product recovery scrubber								
A	2.93×10^{1}	96.4%	2.93×10^{1}	99.6%	2.42×10^{1}	97.6%	2.42×10^{1}	99.6%
Incinerator vent								
B	2.82×10^{-1}	20.8%	2.82×10^{-1}	72.4%	2.43×10^{-1}	28.3%	2.43×10^{-1}	79.9%
Adsorber system vent A								
C	2.60×10^{-1}	16.3%	2.60×10^{-1}	41.4%	1.87×10^{-1}	18.9%	1.87×10^{-1}	42.9%
Adsorber system vent B								
C	2.60×10^{-1}	16.3%	2.60×10^{-1}	41.4%	1.87×10^{-1}	18.9%	1.87×10^{-1}	42.9%
Adsorption system vent								
D	9.11×10^{-1}	45.9%	9.11×10^{-1}	89.4%	7.40×10^{-1}	54.8%	7.40×10^{-1}	92.5%
TOTALS								
A	3.04×10^{1}		2.94×10^{1}		2.48×10^{1}		2.43×10^{1}	
B	1.35		3.90×10^{-1}		8.58×10^{-1}		3.04×10^{-1}	
C	1.60		6.28×10^{-1}		9.89×10^{-1}		4.35×10^{-1}	
D	1.98		1.02		1.37		8.1×10^{-1}	

(continued)

TABLE 4-3. (continued)

	1.5 km				2.0 km			
	No fugitive or storage controls		Fugitive and storage controls[a,b]		No fugitive or storage controls		Fugitive and storage controls[a,b]	
Source/alternative	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]	Concentration[d] μg/m³	% of total emissions[c]
Storage tank A								
A	1.87×10^{-1}	1.1%	1.87×10^{-2}	0.11	1.18×10^{-1}	0.9%	1.18×10^{-2}	0.09%
B	1.87×10^{-1}	37.0%	1.87×10^{-2}	8.8%	1.18×10^{-1}	34.1%	1.18×10^{-2}	7.3%
C	1.87×10^{-1}	33.2%	1.87×10^{-2}	6.9%	1.18×10^{-1}	31.6%	1.18×10^{-2}	6.2%
D	1.87×10^{-1}	22.1%	1.87×10^{-2}	3.4%	1.18×10^{-1}	20.0%	1.18×10^{-2}	2.9%
Storage tank B								
A	2.54×10^{-2}	0.15%	2.54×10^{-3}	0.01%	1.59×10^{-2}	0.12%	1.59×10^{-3}	0.01%
B	2.54×10^{-2}	5.0%	2.54×10^{-3}	1.2%	1.59×10^{-2}	4.6%	1.59×10^{-3}	1.0%
C	2.54×10^{-2}	4.5%	2.54×10^{-3}	0.9%	1.59×10^{-2}	4.3%	1.59×10^{-3}	0.8%
D	2.54×10^{-2}	3.0%	2.54×10^{-3}	0.5%	1.59×10^{-2}	2.7%	1.59×10^{-3}	0.4%
Fugitive sources								
A	1.12×10^{-1}	0.64%	1.12×10^{-2}	0.07%	7.02×10^{-2}	0.54%	7.02×10^{-3}	0.06%
B	1.12×10^{-1}	22.2%	1.12×10^{-2}	5.3%	7.02×10^{-2}	20.3%	7.02×10^{-3}	4.3%
C	1.12×10^{-1}	19.9%	1.12×10^{-2}	4.1%	7.02×10^{-2}	18.8%	7.02×10^{-3}	3.7%
D	1.12×10^{-1}	13.2%	1.12×10^{-2}	2.0%	7.02×10^{-2}	11.9%	7.02×10^{-3}	1.7%
Product recovery scrubber								
A	1.71×10^{1}	98.3%	1.71×10^{1}	99.8%	1.27×10^{1}	98.5%	1.27×10^{1}	99.9%
Incinerator vent								
B	1.80×10^{-1}	35.7%	1.80×10^{-1}	84.7%	1.42×10^{-1}	41.0%	1.42×10^{-1}	87.4%
Adsorber system vent A								
C	1.20×10^{-1}	21.3%	1.20×10^{-1}	44.1%	8.44×10^{-2}	22.6%	8.44×10^{-2}	44.6%
Adsorber system vent B								
C	1.20×10^{-1}	21.3%	1.20×10^{-1}	44.1%	8.44×10^{-2}	22.6%	8.44×10^{-2}	44.6%
Adsorption system vent								
D	5.24×10^{-1}	61.8%	5.24×10^{-1}	94.2%	3.85×10^{-1}	65.4%	3.85×10^{-1}	95.1%
TOTALS								
A	1.74×10^{1}		1.71×10^{1}		1.29×10^{1}		1.27×10^{1}	
B	5.05×10^{-1}		2.12×10^{-1}		3.46×10^{-1}		1.62×10^{-1}	
C	5.64×10^{-1}		2.72×10^{-1}		3.73×10^{-1}		1.89×10^{-1}	
D	8.48×10^{-1}		5.56×10^{-1}		5.89×10^{-1}		4.05×10^{-1}	

(continued)

TABLE 4-3. (continued)

	10.0 km				20.0 km			
	No fugitive or storage controls		Fugitive and storage controls[a,b]		No fugitive or storage controls		Fugitive and storage controls[a,b]	
Source/alternative	Concentration[d] $\mu g/m^3$	% of total emissions[c]	Concentration[d] $\mu g/m^3$	% of total emissions[c]	Concentration[d] $\mu g/m^3$	% of total emissions[c]	Concentration[d] $\mu g/m^3$	% of total emissions[c]
Storage tank A								
A	8.79×10^{-3}	0.63%	8.79×10^{-4}	0.06%	3.05×10^{-3}	0.61%	3.05×10^{-4}	0.06%
B	8.79×10^{-3}	23.8%	8.79×10^{-4}	3.4%	3.05×10^{-3}	27.5%	3.05×10^{-4}	4.8%
C	8.79×10^{-3}	28.1%	8.79×10^{-4}	4.9%	3.05×10^{-3}	27.7%	3.05×10^{-4}	4.8%
D	8.79×10^{-3}	15.6%	8.79×10^{-4}	2.1%	3.05×10^{-3}	15.1%	3.05×10^{-4}	2.0%
Storage tank B								
A	1.14×10^{-3}	0.08%	1.14×10^{-4}	0.01%	3.97×10^{-4}	0.08%	3.97×10^{-5}	0.01%
B	1.14×10^{-3}	3.1%	1.14×10^{-4}	0.49%	3.97×10^{-4}	3.6%	3.97×10^{-5}	0.62%
C	1.14×10^{-3}	3.6%	1.14×10^{-4}	0.64%	3.97×10^{-4}	3.6%	3.97×10^{-5}	0.63%
D	1.14×10^{-3}	2.0%	1.14×10^{-4}	0.27%	3.97×10^{-4}	2.0%	3.97×10^{-5}	0.26%
Fugitive sources								
A	5.05×10^{-3}	0.36%	5.05×10^{-4}	0.04%	1.75×10^{-3}	0.35%	1.75×10^{-4}	0.04%
B	5.05×10^{-3}	13.7%	5.05×10^{-4}	2.2%	1.75×10^{-3}	15.6%	1.75×10^{-4}	2.7%
C	5.05×10^{-3}	16.1%	5.05×10^{-4}	2.8%	1.75×10^{-3}	15.9%	1.75×10^{-4}	2.8%
D	5.05×10^{-3}	9.0%	5.05×10^{-4}	1.2%	1.75×10^{-3}	8.7%	1.75×10^{-4}	1.1%
Product recovery scrubber								
A	1.37	98.6%	1.37	99.9%	4.98×10^{-1}	99.0%	4.98×10^{-1}	99.9%
Incinerator vent								
B	2.19×10^{-2}	59.4%	2.19×10^{-2}	93.6%	5.89×10^{-3}	53.1%	5.89×10^{-3}	91.9%
Adsorber system vent A								
C	8.14×10^{-3}	26.0%	8.14×10^{-3}	45.8%	2.91×10^{-3}	26.4%	2.91×10^{-3}	45.9%
Adsorber system vent B								
C	8.14×10^{-3}	26.0%	8.14×10^{-3}	45.8%	2.91×10^{-3}	26.4%	2.91×10^{-3}	45.9%
Adsorption system vent								
D	4.13×10^{-2}	73.4%	4.13×10^{-2}	96.5%	1.50×10^{-2}	74.4%	1.50×10^{-2}	96.6%
TOTALS								
A	1.39		1.38		5.03×10^{-1}		4.98×10^{-1}	
B	3.69×10^{-2}		2.34×10^{-2}		1.11×10^{-2}		6.41×10^{-3}	
C	3.13×10^{-2}		1.78×10^{-2}		1.10×10^{-2}		6.34×10^{-3}	
D	5.63×10^{-2}		4.28×10^{-2}		2.02×10^{-2}		1.55×10^{-2}	

Notes for Table 4-3.

NOTE: Data adjusted for a benzene conversion rate of 90 percent.

[a] Storage tanks A and B are assumed to be controlled to the 90-percent level.

[b] Fugitive sources are assumed to be controlled to the 90-percent level.

[c] Percentage of total emissions is the percentage of the total emitted for that particular source and alternative combination.

[d] Concentration ppmv = (0.000314) × concentration in $\mu g/m^3$ (at 25° C and at 1 atm).

SOURCE: H. E. Cramer Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions from a Maleic Anhydride Plant for Four Emission Control Options. Salt Lake City, Utah. (Prepared for Source-Receptor Analysis Branch, U.S. Environmental Protection Agency. Research Triangle Park, N.C.). EPA Contract No. 68-02-2507. August 1978.

calculated the maximum benzene concentrations from the combined emissions for averaging times of 1 hour, 3 hours (uncontrolled case only), 8 hours, and 24 hours. These data are shown in Table 4-4.

The short-term program (ISCST) assessed the averaging times of 1,3,8, and 24 hours. The long-term program (ISCLT) was used to calculate annual averages. For the model features used in this study, the ISCST program corresponds to the Single-Source (CRSTER) Model, modified to include the effects of separation of individual sources and the effects of volume and area source emissions. The ISCLT program, which is a sector-averaged model similar to the Air Quality Display Model (AQDM) or the Climatological Dispersion Model (CDM), makes the same basic model assumptions as the ISCST program. The ISCLT program uses STAR summaries (statistical tabulation of the joint frequency of occurrence of wind-speed and wind-direction categories, classified according to the Pasquill stability categories) to calculate annual average concentrations. For the same source data and meteorological data base, ISCLT calculates annual average concentrations that are equivalent to those obtained from ISCST using a year of sequential, hourly meteorological data.

The dispersion estimates used a model plant with a capacity of 22,700 Mg/yr. The layout of the emission points was assumed to be as depicted in Figure 4-1. The emission points are identified by name and number in Table 4-5. Fugitive emissions and emissions from benzene storage were assumed to be uncontrolled, and estimated emission rates (at design capacity) were taken from Reference 3; these were 0.45 gm/sec (9.9×10^{-4} lb/sec), 0.05 gm/sec (1.1×10^{-4} lb/sec), and 0.22 gm/sec (4.8×10^{-4} lb/sec) for Tank A, Tank B, and fugitive emissions, respectively. These types of emissions will be covered in separate documents but were included here to illustrate the relative influence of the product recovery absorber and refining vacuum system vents on the ambient benzene concentration.

Two maleic anhydride plants are located in the Pittsburgh, Pennsylvania, area. Also, meteorological conditions that maximize ground level concentrations produced by emissions from stacks of the type studied here (neutral stability in combination with moderate-to-strong winds that persist within a narrow angular sector for a number of hours) are common in the Pittsburgh area. Surface-air and upper-air meteorological data from the Greater

TABLE 4-4. MAXIMUM 1-HOUR, 3-HOUR, 8-HOUR, 24-HOUR, AND ANNUAL AVERAGE GROUND LEVEL BENZENE CONCENTRATIONS PRODUCED BY THE COMBINED EMISSIONS FROM A MALEIC ANHYDRIDE PLANT AT ANY DISTANCE DOWNWIND AND AT 0.1, 1.0, 10.0, AND 20.0 km

	Concentration (μg/m^3)[a]			
Distance (km)	Alternative A	Alternative B	Alternative C	Alternative D
1-hour concentrations				
0.3	1.31×10^3	-	-	-
0.1	1.11×10^3	3.73×10^2	3.73×10^2	4.45×10^2
1.0	9.93×10^2	4.32×10^1	5.23×10^1	6.31×10^1
10.0	1.46×10^2	3.28	3.26	6.03
20.0	7.63×10^1	1.71	1.70	3.11
3-hour concentrations				
0.7	9.68×10^2	-	-	-
0.1	2.80×10^2	-	-	-
1.0	8.14×10^2	-	-	-
10.0	7.28×10^1	-	-	-
20.0	2.90×10^1	-	-	-
8-hour concentrations				
0.3	6.48×10^2	-	-	-
0.1	3.55×10^2	1.55×10^2	1.56×10^2	1.62×10^2
1.0	4.41×10^2	1.34×10^1	1.79×10^1	2.50×10^1
10.0	3.92×10^1	8.68×10^{-1}	9.05×10^{-1}	1.62
20.0	1.48×10^1	3.29×10^{-1}	2.55×10^{-1}	6.01×10^{-1}
24-hour concentrations				
0.3	5.82×10^2	-	-	-
0.1	1.95×10^2	8.83×10^1	8.83×10^1	1.02×10^2
1.0	3.08×10^2	6.09	7.24	1.27×10^1
10.0	1.52×10^1	3.37×10^{-1}	3.52×10^{-1}	6.31×10^{-1}
20.0	5.79	1.28×10^{-1}	1.28×10^{-1}	2.34×10^{-1}
Annual concentrations				
0.3	6.10×10^1	-	-	-
0.1	2.16×10^1	2.07×10^1	2.07×10^1	2.07×10^1
1.0	3.90×10^1	1.26	1.48	2.03
10.0	2.18	4.89×10^{-1}	4.94×10^{-2}	8.89×10^{-2}
20.0	7.92×10^{-1}	1.75×10^{-2}	1.74×10^{-2}	3.18×10^{-2}

[a] Concentration in ppmv = (0.000314) × concentration in μg/m^3 (at 25° C [77° F] and 1 atm)..

SOURCE: H. E. Cramer, Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions From a Maleic Anhydride Plant for Four Emission Control Options. Salt Lake City, Utah. (Prepared for Source-Receptor Analysis Branch, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2507. August 1978.

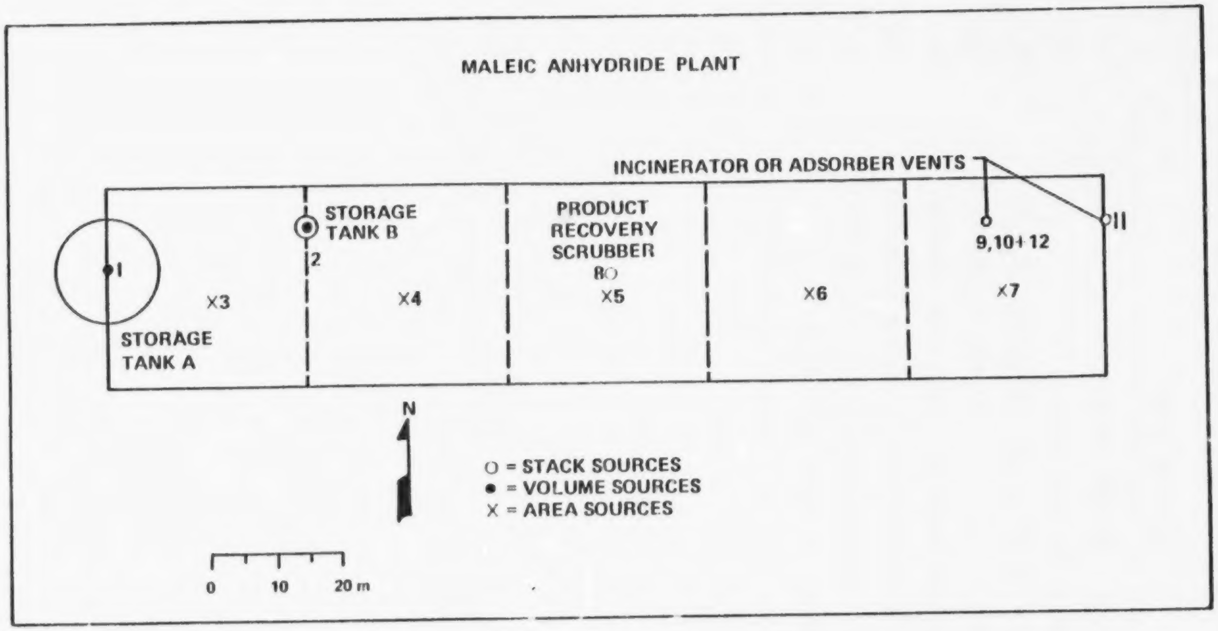
MALEIC ANHYDRIDE PLANT
INCINERATOR OR ADSORBER VENTS
1
STORAGE TANK A
X3
2
STORAGE TANK B
X4
PRODUCT RECOVERY SCRUBBER
8
X5
X6
9,10+12
11
X7
N
O = STACK SOURCES
● = VOLUME SOURCES
X = AREA SOURCES
0
10
20 m

Figure 4-1. Layout of the maleic anhydride plant showing source locations.

TABLE 4-5. IDENTIFICATION OF MODEL SOURCE NUMBERS BY SOURCE NAME AND EMISSION CONTROL ALTERNATIVE

Model source no.	Source name	Emission control alternatives
1	Benzene storage tank A	All
2	Benzene storage tank B	All
3	Fugitive emissions (subarea source)	All
4	Fugitive emissions (subarea source)	All
5	Fugitive emissions (subarea source)	All
6	Fugitive emissions (subarea source)	All
7	Fugitive emissions (subarea source)	All
8	Product recovery scrubber	A--uncontrolled
9	Incinerator vent (with heat recovery)	B--99 percent
10	Adsorber system vent A	C--99 percent
11	Adsorber system vent B	C--99 percent
12	Adsorber system vent	D--97 percent

SOURCE: H. E. Cramer Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions From a Maleic Anhydride Plant for Four Emission Control Options. Salt Lake City, Utah. (Prepared for Source-Receptor Analysis Branch, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2507. August 1978.

Pittsburgh Airport were used in the dispersion-model calculations for these reasons and because the data were readily available. The maleic anhydride plant was assumed to be located in the Pittsburgh urban area, and ISC Urban Mode 2 was used in the dispersion-model calculations to account for the effects of urban roughness elements and heat sources. Because of the high frequency of occurrence of west winds in the Pittsburgh area, an east-west orientation was assumed for the maleic anhydride plant (see Figure 4-1).

The ISC Model contains numerous model options and features that are required to simulate the air quality impact of emissions from the wide variety of sources found in industrial source complexes. Because of the requisite complexity of the ISC Model, it is usually economically unfeasible to use the ISCST program to calculate hourly concentrations for a year of sequential, hourly meteorological data--the procedure generally followed with the Single-Source (CRSTER) Model. However, with knowledge of the critical meteorological conditions for emissions from maleic anhydride plants, it was possible to select ten 24-hour periods representative of "worst-case" dispersion conditions and to execute ISCST for these days only to obtain maximum 1-hour, 3-hour (Alternative A only), 8-hour, and 24-hour average benzene concentrations. First, the ratio of the vector mean wind speed to the scalar mean wind speed was computed for each day of 1964 from the Greater Pittsburgh Airport hourly surface observations. This ratio has a value near unity for days of persistent wind directions that are usually associated with maximum 8-hour and 24-hour average concentrations. The days with ratio values near unity were examined to select 10 days with neutral and/or stable conditions and either persistent light-to-moderate winds ("worst-case" meteorological conditions for the storage tanks and fugitive emissions) or persistent moderate-to-strong winds ("worst-case" meteorological conditions for the stack emissions). Those selected were days 5, 10, 26, 112, 118, 122, 265, 316, 343, and 358. The principal meteorological input to the ISCLT program for the annual concentration calculations was the 1964 annual STAR summary for the Greater Pittsburgh Airport.

The objective of the dispersion study was to estimate maximum ground level benzene concentrations at or beyond the property boundaries and at downwind distances of 0.1, 1.0, 10.0, and 20.0 km (0.06, 0.62, 6.21, and 12.43 mi). Maximum ground level concentrations produced by fugitive and storage tank emissions can be expected to occur in the immediate vicinity of the plant production area. However, the dispersion coefficients used by the ISC Model, the Pasquill-Gifford curves, begin at a downwind distance of 100 m (109.4 yd). Consequently, the property boundaries were assumed to be 100 m (109.4 yd) from the edge of the plant production area, and receptors were spaced at 30-m (32.8-yd) intervals around the edge of the property boundaries. Preliminary calculations for the buoyant stack emissions indicated that the maximum ground level concentrations should occur within 2 km of the stacks. Therefore, additional receptors were placed at distances from the boundaries, 0.1., 0.2, 0.4, 0.6, 0.9, 1.4, 1.9, 9.9, and 19.9 km (0.06, 0.12, 0.25, 0.37, 0.56, 0.87, 1.18, 6.15, and 12.37 mi) along radials originating at the center of Source 5 and passing through each of the receptors at the property boundary.

The results of the dispersion analysis are summarized in Tables 4-1, 4-2, 4-3, and 4-4. The ISC Model calculations for the uncontrolled case, Alternative A, indicate that emissions from the product recovery absorber and refining vacuum system vent are principally responsible for the maximum ground level benzene concentrations produced by the combined emissions from the plant. At distances where maximum concentrations occur, the product recovery absorber emissions account for more than 85 percent of the calculated concentrations. Additionally, for some distances and averaging times, the product recovery absorber accounts for more than 98 percent of the calculated maximum concentrations. However, the product recovery absorber of Alternative A does not necessarily dominate the maximum concentrations at the assumed boundaries of the plant property. For a given averaging time, meteorological conditions such as stability and wind direction determine whether the fugitive and storage tank emissions or the product recovery absorber emissions are primarily responsible for the magnitude and location of the maximum concentration at the property boundaries.

For Alternatives B, C, and D, the benzene emissions from the product recovery absorber are reduced by 97 to 99 percent. With these reduced emissions, the ISC Model calculations show that the fugitive and storage tank emissions (which are not controlled) wield a much greater influence on the maximum ground level benzene concentrations than in Alternative A. The maximum concentrations calculated for all averaging periods are located at the assumed boundaries of the plant property, and the fugitive and storage tank emissions are principally responsible for the maximum concentrations. Because the fugitive and storage tank emissions dominate the air quality near the plant for Alternatives B, C, and D, the effects on ambient air quality near the plant of different emission reductions are minimal without control of fugitive and storage emissions. Disregarding the contribution from fugitive and storage emissions, however, Table 4-1 shows that the maximum concentration of benzene from the process vents after 99 percent control is less than 0.1 percent of the ambient concentration resulting from these sources when uncontrolled. When Alternative D is used (97 percent control by carbon adsorption), the resulting benzene concentration is about 0.1 percent of the concentration before control.

Table 4-1 and Table 4-4 emphasize the positive impact on air quality from application of the alternative control systems. Table 4-6 summarizes the reduction in benzene emissions for both a 97-percent and 99-percent control level on a plant-by-plant basis, using a benzene conversion rate of 90 percent. However, some adverse effects on air quality are usually associated with each control technique incorporated in the regulatory options. These adverse impacts are compared to the benefits for each technique in the next sections.

4.1.1 Carbon Adsorption (99 Percent Control of Benzene)

The carbon adsorption system described for this level of control in Chapter 2 will reduce benzene emissions from 1,540 Mg/yr (1,700 tons/yr) to 50 Mg/yr (55 tons/yr) for the model plant. Of this emission reduction, about 1,450 Mg (1,600 tons) of benzene per year are recovered for recycling to the process.[3] The remaining benzene and the other VOC (~460 Mg/yr [~507 tons/yr]) are picked up in the scrubber or decanter water and are removed as effluent wastewater.[3] The amount of benzene sent to wastewater treatment is approximately 40 Mg/yr (44 tons/yr), or about 1.4 g/sec (3.1×10^{-3} lb/sec).

All plants should be able to recycle the benzene-containing wastewaters as makeup water for the absorbers.[4] Wastewaters typically include those from water of dehydration, water removed from scrubber solution, and the condensed vacuum jet stream.[3] If wastewaters from the carbon adsorption units are recycled, some plants will have to discharge previously

TABLE 4-6. REDUCTION IN BENZENE EMISSIONS FOR SELECTED CONTROL LEVELS[1]

	Capacity (Mg/yr)	Total uncontrolled emissions[2] (Mg/yr)	Existing control device efficiency (%)	Emissions with existing control[3] (Mg/yr)	97% control		99% control	
					Allowable benzene emissions[4] (Mg/yr)	Incremental benzene removed[5] (Mg/yr)	Allowable benzene emissions[6] (Mg/yr)	Incremental benzene removed[7] (Mg/yr)
Ashland, Neal W. Va.	27,200	1,849	0	1,849	126	1,723	61	65
Koppers, Bridgeville, Pa.	15,400	1,047	99	26	72	0	35	0
Monsanto, St. Louis, Mo.	38,100[12]	2,589	0	2,589	177	2,412	85	92
DENKA, Houston, Tex.	22,700	1,543	97	68	105	0	50	18
Reichhold, Elizabeth, N.J.	13,600	924	97	41	63	0	30	11
Reichhold, Morris, Ill.	20,000	1,359	90	154	93	61	45	48
Tenneco, Fords, N.J.	11,800	802	0	802	55	747	26	29
U.S. Steel, Neville Island, Pa.	38,500	2,616	90	295	179	116	86	93
Totals		12,729		5,824 Mg/yr	870 Mg/yr	5;059 Mg/yr	420 Mg/yr	356 Mg/yr

[1] Emission rates are based on: Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

[2] Total uncontrolled emissions = [(2.6 kg/hr × 8,760 hr) + (190 kg/hr × 8,000 hr)]

$$\times \frac{\text{production capacity Mg/yr}}{\text{model plant capacity 22,700 Mg/yr}} .$$

Based on a benzene conversion rate of 94.5 percent.

[3] Emissions with existing control = {(2.6 kg/hr × 8,760 hr) + [190 kg/hr × (1-control device efficiency)

$$\times \text{ 8,000 hr]}\} \times \frac{\text{production capacity Mg/yr}}{\text{model plant capacity 22,700 Mg/yr}} .$$

[4] Allowable benzene emissions with 97 percent control = {[(2.6 kg/hr × 8,760 hr) + (345.4 kg/hr × 0.03 × 8,000 hr)] × (production capacity/22,700 Mg/yr)}. Based on a benzene conversion rate of 90 percent.

[5] Incremental benzene removed with 97 percent control = (emissions with existing controls) - (allowable benzene emissions at 97% control).

[6] Allowable benzene emissions with 99 percent control = {[(2.6 kg/hr × 8,760 hr) + (345.4 kg/hr × 0.01 × 8,000 hr)] × (production capacity/22,700 Mg/yr)} Based on a benzene conversion rate of 90 percent.

[7] Incremental benzene removed with 99 percent control = (emissions with 97% controls) - (allowable benzene emissions at 99% control).

recycled wastewaters either to the plant's wastewater treatment plant or to a municipal wastewater treatment plant. These wastewaters discharged to treatment facilities will have lower benzene concentrations than the waste stream from the carbon adsorbers but will contain additional organic compounds; e.g., maleic acid.[5]

Depending upon the method of wastewater treatment practiced by the facility, some volatilization of benzene to the atmosphere can occur. A worst-case assumption is that all of this benzene is emitted to the atmosphere at a constant rate of 1.4 g/sec (3.1×10^{-3} lb/sec). This rate is roughly 3 percent of the benzene emission rate from the uncontrolled product recovery absorber of a model plant, which was shown in Table 4-1. The actual quantity of benzene emitted has not been determined because it is highly dependent upon the wastewater treatment technique because benzene is biodegradable.[6] Furthermore, at least two facilities employ an aqueous waste incinerator for the maleic anhydride production facilities.[7,8] The aqueous stream from the carbon adsorber system could conceivably be combined with the other liquid wastes sent to the incinerator. Assuming a conservative benzene destruction efficiency of 90 percent in this type of incinerator, the benzene emissions from this source would be about 0.14 g/sec (3.1×10^{-4} lb/sec), or less than 1 percent of the emissions from an uncontrolled product recovery absorber.

4.1.2 Carbon Adsorption (97 Percent Control of Benzene)

The carbon adsorption system described for this level of control will reduce benzene emissions from 1,540 Mg/yr (1,700 tons/yr) to 100 Mg/yr (110 tons/yr) based on the model plant. Like the system discussed in Section 4.1.1, some fraction of this controlled benzene is picked up in the decanter water and removed as effluent wastewater. In a manner analogous to the previous system, about 1,430 Mg (1,580 tons) of benzene per year are recovered and recycled to the process. Therefore, about 40 Mg/yr or 44 tons/yr (~1.4 g/sec [3.1×10^{-3} lb/sec]) of benzene would be carried in the aqueous waste from the adsorber system.

Consistent with the assumptions in Section 4.1.1, a worst-case benzene emission rate from this source is 1.4 g/sec (3.1×10^{-3} lb/sec), or less than 3 percent of the emissions from an uncontrolled product recovery absorber. Similarly, control in a liquid incinerator would result in a benzene emission of less than 0.14 g/sec (3.1 × 10^{-4} lb/sec). Recycle of the aqueous stream could also be practiced.

4.1.3 Thermal Incineration (99 Percent Control of Benzene)

The incinerator system described in Chapter 2 reduces benzene emissions from 1,540 Mg/yr (1,700 tons/yr) to 50 Mg/yr (55 tons/yr) for the model plant. No organics are recovered for recycle. Pollutants generated by the combustion process, particularly nitrogen oxides (NO_x), may have a negative impact on the environment. Compared with industrial furnaces and boilers, however, fume incinerators tend to have low NO_x emission factors. One reference reports NO_x effluent concentrations from thermal incinerators of 20 to 30 ppmv.[9] Even at temperatures as high as 870° C (1,598° F), estimated NO_x concentrations would be on the order of 50 to 200 ppmv. Preliminary results from recent tests on an incinerator operating at 760° C (1,400° F) showed an average NO_x emission rate of about 0.27 g/sec (6.0 × 10^{-4} lb/sec).[10]

The waste gas stream going to the incinerator contains approximately 2 percent CO by volume. Because of sufficient air and an incinerator temperature of 870° C (1,598° F) (this exceeds the autoignition temperature of CO, which is 651° C [1,204° F]), essentially all CO is oxidized. The incinerator thus achieves a net reduction in CO emissions from maleic anhydride plants. For the incinerator conditions described in Chapter 2, it is estimated that greater than 99 percent control of CO would be achieved.[3]

In locations where natural gas may be unavailable for the incinerator, fuel oil will be used as supplemental fuel. Depending upon the type of fuel used, there may be SO_2 and particulate emissions from this control technique. It is predicted that 87,000 bbl of oil per year will be used to attain 99 percent control by thermal incineration. If this oil were 0.3 percent sulfur by weight, 64 Mg of SO_2 would be released to the atmosphere per year.

4.1.4 Thermal Incineration (97 Percent Control of Benzene)

An incinerator with this efficiency will reduce benzene emissions from 1,540 Mg/yr (1,700 tons/yr) to 100 Mg/yr (110 tons/yr) for the model plant. As with the incinerator discussed in the previous section, negative environmental impacts may result from NO_x, SO_x, and particulate matter emissions, but these impacts would tend to be less than impacts of the incinerator operated at a higher temperature. It is predicted that 29,000 bbl of oil per year will be used to attain 97 percent control by thermal incineration. If the oil were 0.3 percent sulfur by weight, 26 Mg of SO_2 would be released to the atmosphere per year. The efficiency of CO destruction would also be reduced at the lower temperature of 760° C (1,400° F).

4.1.5 n-Butane Process Conversion (100 Percent Control of Benzene)

Although few data are yet available on the quantity and composition of the emissions, preliminary information indicates that total uncontrolled VOC emissions are higher in the n-butane than in the benzene oxidation process.[8] n-Butane used as a feedstock may result in increased photochemical smog; it has been found to be photochemically oxidized to peroxides, which are the precursors of the various types of photochemical smog.[11,12] At present, there is no nationwide requirement to control the VOC emissions from the n-butane oxidation process for maleic anhydride production. Future regulations will be prepared for emissions from oxidation systems, such as n-butane. Trace quantities of benzene may be formed in the n-butane oxidation process.[13,14] However, one company using n-butane has found no benzene emissions from its reactors at lower detection limits of 1 ppmv.[15]

If benzene is not used for new maleic anhydride production, benzene supply intended for that production could be added to gasoline as an octane booster. Total benzene emissions from gasoline marketing would then increase. Several factors make this scenario seem unlikely, however.

Currently, benzene demand often exceeds the available supply, a situation expected to continue through the 1980's. Consequently, the benzene that would be used at new sources, if there were no standard, is not part of current production output. If new maleic anhydride sources wanted to use benzene as their feedstock, an increase in benzene demand would result.

Typically, when demand for benzene fluctuates, supply is adjusted by changing the level of production from the most expensive source. If benzene were not prohibited as a feedstock for new sources and there were little, if any, slack in the benzene supply, the additional benzene required would probably be supplied by hydrodeakylation (HDA). HDA is the most expensive benzene production method, and changes in benzene demand can be accommodated by changing the volume of benzene production from HDA. HDA production currently accounts for 25 to 30 percent of the benzene produced per year.[16] Because existing maleic anhydride plants currently use 3 percent of the benzene produced, the HDA process should be able to accommodate fluctuations in demand for benzene caused by maleic anhydride producers.

If chemical process industries know that benzene will not be used as feedstock for maleic anhydride production, they will presumably adjust their projections of future benzene demand. Additional benzene production capacity would be adjusted to reflect this decrease in demand for benzene as a feedstock for new maleic anhydride sources.

In addition, toluene, the feedstock for benzene in the HDA process, is a better octane booster than benzene. Toluene's octane number ([research plus motor]/2) is 102.9[17] as compared to 97 for benzene.[18] Unleaded premium has an octane number of 93.0.[17] Thus, instead of producing benzene by HDA and blending it into gasoline, it seems more logical to blend toluene into the gasoline than to produce more benzene.

4.2 WATER POLLUTION IMPACT

No water effluents are discharged as a result of the application of incinerator systems. A wastewater stream containing benzene is associated with the carbon adsorption systems. However, the organic load of the wastewater from carbon adsorption would be less than 10 percent of the total liquid waste from a typical maleic anhydride production facility.[3,19] The wastewater stream from the carbon adsorption systems could also be treated along with the other plant effluent or recycled to the process. The organic liquid effluent from the carbon adsorber systems will include maleic acid, other organic acids, formaldehyde, other aldehydes, and benzene. These materials are also contained in other waste liquid streams from the process. The organic liquid effluent from the carbon adsorption systems is a small percentage of the total liquid effluent from maleic anhydride production facilities and is therefore estimated to have an insignificant incremental impact on water pollution.

Insufficient data are available to assess the wastewater load from the n-butane process relative to the load from the benzene process. The difference in wastewater load will depend upon the specific operation of the recovery, dehydration, and refining steps. As discussed in Chapter 1, differences between the processes are not well known. Catalysts presently used in the n-butane conversion process produce a greater quantity of water-soluble byproducts than catalysts used in the benzene conversion process.[13] The latest catalysts developed yield liquid and solid wastes of the same order of magnitude as catalysts from the benzene process.[20] Formation of small quantities of fumaric acid in dehydration cannot be avoided. Formic acid and maleic anhydride are the bases for some powerful adhesives, which could lead to plugged pipes and equipment. To prevent this plugging, periodic water washes of equipment are necessary, possibly leading to an impact on water pollution, but no greater than that from the benzene process.[19]

4.3 SOLID WASTE DISPOSAL IMPACT

One potential impact on solid waste disposal associated with the alternative emission control systems is the handling of spent carbon from the adsorption systems. Typically, rather than being disposed of in a landfill, the spent carbon will be transported to a facility for reclamation and regeneration. If it were disposed of in a landfill, the amount of solid waste from this source would be approximately 7,620 kg/yr (8.4 tons/yr) from the model plant, using the system achieving 99 percent control of benzene.[3] Assuming the same bed life for the system at 97 percent control of benzene, the amount of solid waste would be on the order of 7,430 kg/yr (8.2 tons/yr) for the model plant. However, in both cases it is likely that the carbon can be reclaimed. Because n-butane conversion to maleic anhydride is not as efficient as benzene conversion, it yields lower recovery of the product as molten maleic anhydride and increased recovery by absorption in water and conversion to maleic acid. This recovery rate not only makes it more difficult to purify the maleic anhydride but requires disposal of increased amounts of acidic organic materials,[13,21] possibly as solid waste. The latest catalyst developed, not currently in use, will yield quantities of liquid and solid wastes of the same order of magnitude as those from the benzene process.[20]

4.4 ENERGY IMPACT

A model uncontrolled maleic anhydride plant, using the benzene process, will produce a small energy surplus of 15 kJ (14.2 Btu) per kilogram of maleic anhydride produced.[3,16] For a model plant with a production capacity of 22,700 Mg/yr, the energy surplus is 340 GJ/yr (3.2×10^8 Btu [55 bbl/yr]). The energy impact of each control technique for a model plant is discussed in the following sections. For comparison, energy equivalents, in barrels of fuel oil, are included in parentheses (assuming 140,000 Btu/gal or 6.2 GJ/bbl).

4.4.1 Carbon Adsorption (99 Percent Control of Benzene)

Energy (steam) is required to desorb the benzene from the carbon. For the model plant, this energy (as steam) is 57,000 MJ/Mg (24,500 Btu/lb) of benzene emission reduced, or 86,000 GJ/yr (8.2×10^{10} Btu/yr).[3] The electrical energy required for recycle pumps and other equipment is 2,500 MJ/Mg (1,080 Btu/lb) of benzene emission reduced, or 3,800 GJ/yr (3.6×10^{9} Btu/yr) for the model plant.[3] The total energy requirement for a model plant is thus about 90,000 GJ/yr (8.5×10^{10} Btu/yr [14,500 bbl/yr]) more than for an uncontrolled plant.

4.4.2 Carbon Adsorption (97 Percent Control of Benzene)

Less energy is required for this system than for the system operating at a higher removal efficiency. For the model plant, the required energy (as steam) is estimated to be 82,000 GJ/yr (7.8×10^{10} Btu/yr), or 56,000 MJ/Mg (24,000 Btu/lb) of benzene emission reduced.[22] The required electrical energy is assumed to be the same as for the system operating at 99 percent control of benzene. Therefore, the total requirement is about 86,000 GJ/yr (8.2×10^{10} Btu/yr [14,000 bbl/yr]) more than for an uncontrolled model plant.

4.4.3 Thermal Incineration (99 Percent Control of Benzene)

Supplemental fuel is required in the form of natural gas or fuel oil to maintain suitable operating conditions. The net amount of energy required for the model plant ranges from 278,000 MJ/Mg (120,000 Btu/lb) of benzene removed for an incinerator without any form of heat recovery to 62,000 MJ/Mg (27,000 Btu/lb) for an incinerator in which 50 percent of the heat in the exit gas stream is recovered.[3] On an annual basis, this corresponds to an additional 420,000 GJ/yr (4.0×10^{11} Btu/yr [68,000 bbl/yr]) for an incinerator without heat recovery and 95,000 GJ/yr (8.9×10^{10} Btu/yr [15,200 bbl/yr]) for an incinerator with heat recovery as compared to an uncontrolled model plant.[3]

4.4.4 Thermal Incineration (97 Percent Control of Benzene)

Because it is expected that 97 percent control of benzene can be achieved at a lower temperature than 99 percent control, the amount of supplemental fuel required will be decreased. For a typical model plant, the expected net energy requirement is about 45,000 GJ/yr (4.1×10^{10}

Btu/yr [7,000 bbl/yr]) for an incinerator with heat recovery, or 29,000 MJ/Mg (12,500 Btu/lb) of benzene emission reduced.[22]

4.4.5 n-Butane Process Conversion

Maleic anhydride recovery from n-butane oxidation may be less than from benzene oxidation, depending on the type of catalyst used. One source states that it presently recovers a 50-percent molar yield of maleic anhydride from benzene oxidation and expects less from n-butane oxidation.[14] Another source claims a 70-percent molar yield from benzene oxidation and a 55-percent molar yield from n-butane oxidation.[21] One recent study places the annualized costs of utilities for the benzene process at $350,000, while annualized utilities costs for the n-butane process are $450,000, based on typical U.S. power and fuel costs.[20] Thus, the n-butane process appears to use more energy than the benzene process.

4.4.6 Summary

All of the add-on control devices require more energy than the typical benzene process produces. The thermal incinerator requires the most energy at 99 percent control, although with heat recovery it requires just slightly more than the carbon adsorption system operating at the same level of control. At 97 percent control, the carbon adsorption system requires about twice as much energy as the incinerator does at 97 percent control. Incineration with heat recovery operating at 99 percent removal efficiency for benzene uses less than 10 percent more energy than the carbon adsorption system operating at 97 percent control of benzene. However, the thermal incinerator operating at 97 percent control uses significantly less energy than the other methods. Furthermore, consideration of the energy usage by the add-on control devices at a 99-percent level of control does not significantly favor either carbon adsorption or incineration, provided heat recovery is employed with the latter. Thermal incineration is favored at 97 percent control. As expected, energy requirement decreases as control efficiency decreases.

The total national energy requirement will depend upon the particular control technique chosen for any alternative level of control. However, the anticipated upper limit on energy usage is the case in which all plants achieve 99 percent control of benzene by thermal incineration with 50 percent heat recovery. The lower limit is the case in which all benzene-based facilities convert to n-butane as the feedstock.

TABLE 4-7. TOTAL NATIONAL INCREMENTAL ENERGY REQUIREMENT

Control system	Annual incremental energy requirement (TJ/yr)
Carbon adsorption	
· 97% control	440 (71,000 bbl/yr)
· 99% control	510 (82,000 bbl/yr)
Thermal incineration (with 50% heat recovery)	
· 97% control	180 (29,000 bbl/yr)
· 99% control	540 (87,000 bbl/yr)
n-Butane process (new facilities only)	
· 100 control	~0 (~0)

Note: Current assumed energy use:

Plant	Control technique	Annual energy requirement (TJ/yr)
DENKA	Thermal incineration	95 (15,200 bbl/yr)
Reichhold, N.J.	Carbon adsorption	52 (8,400 bbl/yr)
Reichhold, Ill.	Carbon adsorption	60 (9,600 bbl/yr)
U.S. Steel	Catalytic incineration	0 (0 bbl/yr)

SOURCE: Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

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Letter from Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Vatavuk, W. M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. July 7, 1978.

TABLE 4-8. n-BUTANE AND BENZENE CHARACTERISTICS

Characteristics	Benzene	Butane
Lower explosive limit	1.3%	1.9%
Upper explosive limit	7.1%	8.5%
Explosion hazard	Moderate	Moderate
Spontaneous heating	No	No
Disaster hazard	Dangerous (highly flammable)	Moderately dangerous (when heated, it emits acrid fumes; can react with oxidizing materials)

SOURCE: Sax, Irving. Dangerous Properties of Industrial Materials, Third Edition. New York, Van Nostrand Reinhold Company, 1968. p. 456, 494.

Table 4-7 reflects the total national energy increase from current control use to that required for a specific type of control and particular control level. Accordingly, plants that currently achieve 97 or 99 percent control are not included in the energy requirements for a 97-percent regulatory option. Also, energy consumed by plants to achieve less than 97 percent control is subtracted from that energy required to achieve 97 percent. Similarly, plants that achieve 99 percent control are not included in the energy requirements for a 99-percent option, and energy used currently to meet any other level of control is subtracted.

4.5 OTHER ENVIRONMENTAL IMPACTS

No other known significant environmental impacts are associated with any of the alternative emission control systems discussed in Chapter 3.

4.6 OTHER ENVIRONMENTAL CONCERNS

4.6.1 Irreversible and Irretrievable Commitment of Resources

Emission control systems using carbon adsorption recover a valuable raw material--benzene--for reuse, whereas an incineration system uses a nonrenewable resource. Over the long term, the amount of recovered benzene is substantial. Moreover, when benzene is recovered rather than burned, there is an energy savings equivalent to the energy required to produce benzene.

4.6.2 Safety Issues

Table 4-8 summarizes the safety-related characteristics of n-butane and benzene. The table shows that if n-butane is substituted for benzene as a feedstock, no significant changes in process safety are expected. n-Butane is already used in similar situations within the petrochemical industry as a feedstock. On a relative basis, n-butane would be safer than benzene.

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5. ECONOMIC IMPACT

5.1 SUMMARY OF ECONOMIC IMPACTS OF THE MALEIC ANHYDRIDE BENZENE NESHAP

Two regulatory options--97 percent and 99 percent--are analyzed for existing facilities. One additional regulatory option--100 percent--is analyzed for new facilities. The following economic issues are examined in this chapter:

- Capital budget impacts,
- Shifts in competitive positions because of unequal control costs,
- Price impacts on products that use maleic anhydride (MA),
- Employment and balance of trade effects,
- Annualized costs in fifth year and energy impacts, and
- Impact of requiring n-butane at new MA facilities.

The results are summarized below.

5.1.1 Capital Budget Impacts

At the 99-percent control level, all firms except probably DENKA and possibly Tenneco would finance the investment. At the 97-percent control level, the only possible closure candidate is Tenneco.

5.1.2 Shifts in Competitive Position

At 97 or 99 percent control and with high levels of demand (100 percent of capacity), no shift in intraindustry competition is expected. At 97 or 99 percent control and with low levels of demand (56 percent), Tenneco would face the highest degree of domestic intraindustry competition, with DENKA, Koppers (Ill.), and Amoco facing the lowest degree of that competition.

5.1.3 Price Impacts On Products That Use MA

The most significant price increase (assuming full cost pass through) projected is in fumaric acid, ranging up to 4.4 percent at the 99-percent control level and low level of MA demand. Polyester resins and malathion will have price increases less than 1 percent under all scenarios.

5.1.4 Employment and Balance of Trade Impacts

At the 99-percent control level--assuming both DENKA and Tenneco close--roughly 45 jobs would be lost. At 97 percent control, if only Tenneco closes, 12 jobs would be eliminated. Under either control level, balance of trade impacts should be negligible. MA is imported only in briquette form, which has a limited market.

5.1.5 Annualized Costs in Fifth Year and Energy Impacts

Annualized costs in the fifth year will be \$2.1 million and \$3.5 million for the 97- and 99-percent control options, respectively. Energy consumption, assuming thermal incineration with 50 percent heat recovery, will be 180 TJ and 540 TJ for the 97- and 99-percent options, respectively.

5.1.6 Impact of Requiring n-Butane (100 Percent Control) at New MA Facilities

Because new MA facilities are expected to be n-butane-based whether or not a standard is promulgated, there is no impact of this requirement.

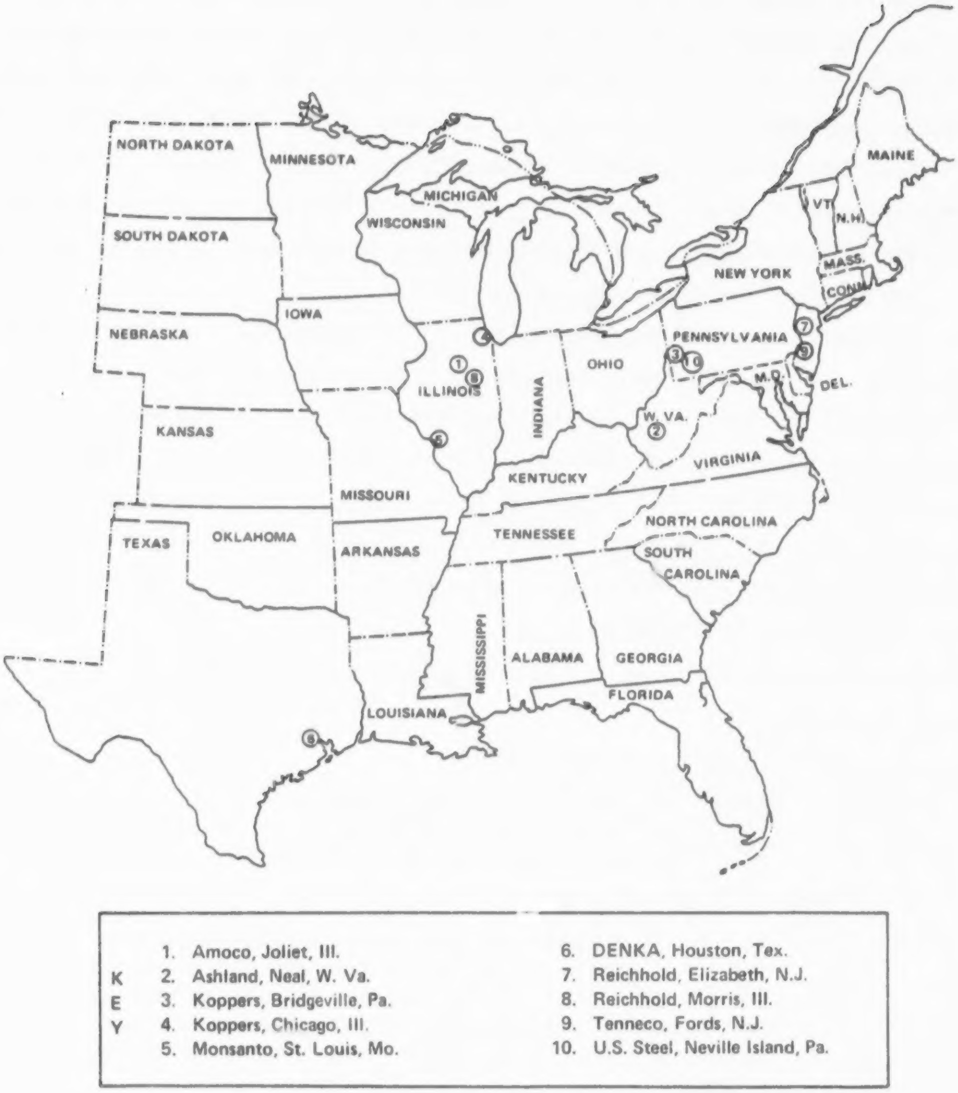
5.2 INDUSTRIAL ECONOMIC PROFILE

5.2.1 Maleic Anhydride Supply and Capacity

5.2.1.1 General. MA is produced by eight companies at 10 plants (Figure 5-1) across 6 States: New Jersey, Pennsylvania, West Virginia, Illinois, Missouri, and Texas. Total employment at these companies is estimated at 330 workers.

Considerable excess MA capacity currently exists in the United States. Market conditions have forced MA producers to operate plants at under 60 percent design capacity. Late in 1978, however, increased demand for MA resulted in improved market conditions. Several companies found it financially viable to increase their plants' operating rates to near full capacity. Although present demand has increased 11 percent since 1976, the predicted demand in 1979 is only 60 percent of the overall name plate capacity of 239 Gg (52.7×10^7 lb).[1]

The overcapacity situation may continue for the next few years. Investor confidence has been strong, however, and producers believe that demand will equal capacity by the end of 1982 (assuming a continued 11-percent growth rate).[2] The confidence felt by chemical producers is best demonstrated by the rapid growth in MA capacity in the past few years. In 1975, annual capacity was rated at 156 Gg (3.4×10^8 lb). In 1976 it grew to 239 Gg (5.3×10^8 lb)--an increase of 53 percent--and included


NORTH DAKOTA
MINNESOTA
MICHIGAN
WISCONSIN
SOUTH DAKOTA
NEBRASKA
IOWA
ILLINOIS
INDIANA
OHIO
PENNSYLVANIA
NEW YORK
MAINE
VT
N.H.
MASS.
CONN
M.D.
DEL.
W. VA.
VIRGINIA
KANSAS
MISSOURI
KENTUCKY
TENNESSEE
NORTH CAROLINA
SOUTH CAROLINA
TEXAS
OKLAHOMA
ARKANSAS
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
LOUISIANA
KEY
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1. Amoco, Joliet, Ill.
2. Ashland, Neal, W. Va.
3. Koppers, Bridgeville, Pa.
4. Koppers, Chicago, Ill.
5. Monsanto, St. Louis, Mo.
6. DENKA, Houston, Tex.
7. Reichhold, Elizabeth, N.J.
8. Reichhold, Morris, Ill.
9. Tenneco, Fords, N.J.
10. U.S. Steel, Neville Island, Pa.

Figure 5-1. Manufacturing locations of maleic anhydride.

SOURCE: Lawson, J.F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride—Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

two new 28-Gg plants (Ashland and Amoco), a large 18.2-Gg (4.0×10^7 lb) expansion (U.S. Steel), and a small 3.6-Gg (7.9×10^6 lb) expansion (Koppers).[3] Furthermore, Monsanto recently announced plans to expand its MA capacity by building a new n-butane-based facility at Pensacola, Florida.[4]

In the United States, MA is produced from one of two feedstocks, benzene and n-butane. At present, benzene supplies are tight and although benzene consumption accounts for 83 percent of all feedstock used in the industry, this number could be reduced significantly should producers convert current benzene-based MA plants to n-butane. Industry sources have indicated that steadily increasing benzene prices have prompted serious consideration of feedstock conversion.[1]

Another source of MA, though minor, is byproduct recovery from phthalic anhydride. Presently, only Koppers in Chicago, with a total capacity of 5 Gg (1.1×10^7 lb), produces MA from the effluent of its phthalic anhydride plant. This source of MA depends on production of phthalic anhydride as the primary product. No growth in this source of MA is expected.

5.2.1.2 The Individual MA-Producing Companies. Table 5-1 compares the capacities of the individual MA plants with their estimated 1978 production, and Figure 5-2 summarizes the relationship between each company's captive and merchant sales. The eight MA-producing companies and their capacities are:

- Amoco Chemical Corporation

  Amoco has the only United States plant totally dedicated to the n-butane process.[5] The facility has an annual capacity of 27 Gg and is expandable to 41 Gg (9.0×10^7 lb).[1]

- Ashland Chemical Company

  The Ashland facility is a new benzene-based plant with an annual capacity of 27 Gg (5.9×10^7 lb) expandable to 41 Gg (9.0×10^7 lb).[5] Fifty percent of the annual capacity is used captively to produce unsaturated polyester resins. This plant can be switched from benzene to n-butane feedstocks.[1]

- Koppers Company, Inc.

  Kopper's Chicago facility can recover 5 Gg (1.1×10^7 lb) of MA per year from the effluent of their phthalic anhydride plant, which started in 1975.[5] The company's Bridgeville plant was mothballed in the spring of 1979. When this 16-Gg/yr (3.5×10^7 lb) benzene plant is operating, approximately 25 percent of the MA produced is used captively to produce unsaturated polyester resins and alkyd resins.

TABLE 5-1. MALEIC ANHYDRIDE CAPACITY

Manufacturing locations	Estimated production[a] 1978 (Gg) (lb)	Capacity 1978 (Gg) (lb)	Process
1. Amoco, Joliet, Ill.	16 (3.5×10^7)	27 (5.9×10^7)	Oxidation of n-butane
2. Ashland, Neal, W.Va.	16 (3.5×10^7)	27 (5.9×10^7)	Oxidation of benzene
3. Koppers, Bridgeville, Pa.	10 (2.2×10^7)	16 (3.5×10^7)	Oxidation of benzene
4. Koppers, Chicago, Ill.	3 (6.6×10^6)	5 (1.1×10^7)	Byproduct of phthalic anhydride manufacture
5. Monsanto, St. Louis, Mo.	29 (5.9×10^7)	48 (1.1×10^8)	(80%) oxidation of benzene (20%) oxidation of n-butane
6. DENKA, USA, Houston, Tex.	14 (3.1×10^7)	23 (5.1×10^7)	Oxidation of benzene
7. Reichhold, Elizabeth, N.J.	11 (2.4×10^7)	18 (4.0×10^7)	Oxidation of benzene
8. Reichhold, Morris, Ill.	16 (3.5×10^7)	27 (6.0×10^7)	Oxidation of benzene
9. Tenneco, Fords, N.J.	7 (1.5×10^7)	12 (2.6×10^7)	Oxidation of benzene
10. U.S. Steel, Neville Island, Pa.	22 (4.9×10^7)	36 (7.9×10^7)	Oxidation of benzene
TOTAL	144 (31.8×10^7)	239 (52.7×10^7)	

[a] Estimated by assuming 56 percent capacity for each plant.

SOURCES: Blackford, J. C. Marketing Research Report on Maleic Anhydride. Chemical Economics Handbook. Stanford Research Inst. Menlo Park, Calif. July 1976.

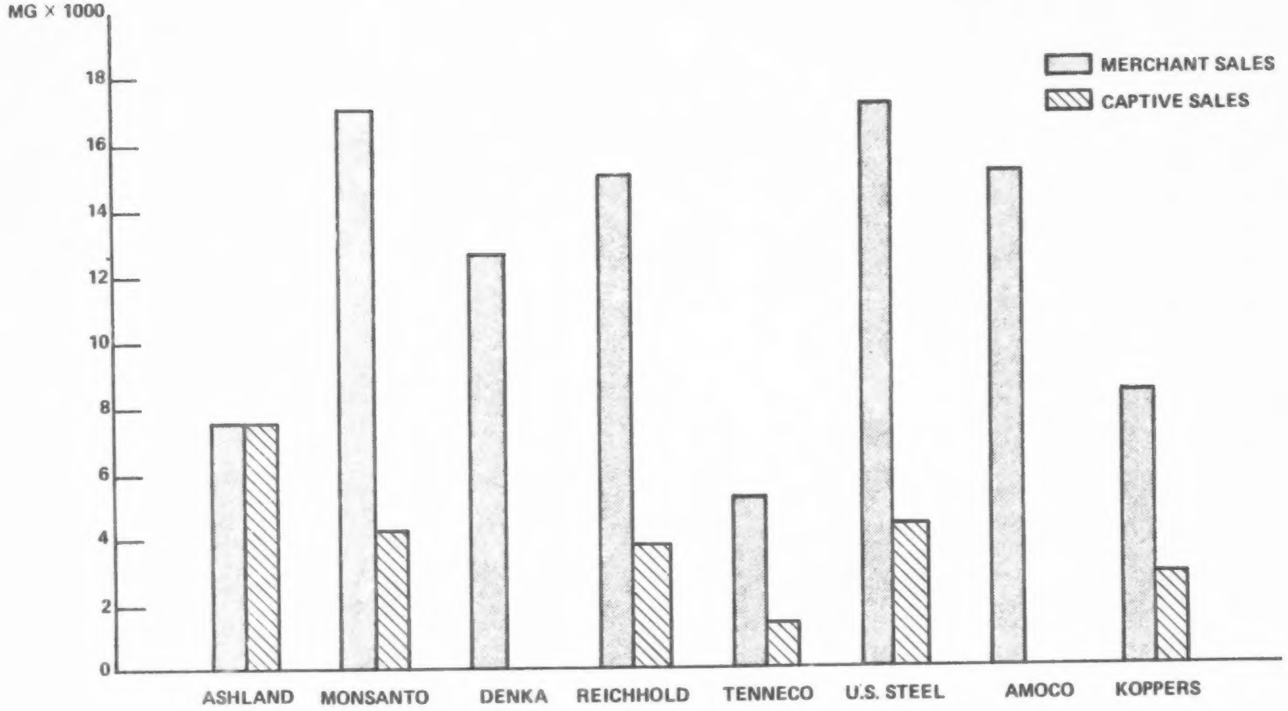

Figure 5-2. Captive and merchant sales of MA companies.

- Monsanto Company

  Monsanto at 48 Gg/yr (1.1×10^8 lb/yr) is the largest producer of MA. The company uses n-butane feedstocks for about 20 percent of its capacity. Approximately 25 percent of the MA is consumed captively to produce fumaric acid, maleate/fumerate esters, styrene copolymers, and ethylene-maleic anhydride copolymers.[5] Monsanto has announced plans to build a new n-butane-based MA facility at Pensacola, Florida.

- DENKA, USA

  Their 23-Gg/yr (5.1×10^7 lb/yr) Houston facility was designed by Scientific Design Company, Inc., and was purchased from Petrotex Chemical Corporation by a Japanese firm, July 1, 1977.[6] The feedstock now used is benzene.

- Reichhold Chemicals, Inc.

  Reichhold's combined production from both plants, at Elizabeth, New Jersey, and Morris, Illinois, is 45 Gg/yr (9.9×10^7 lb/yr), 20 percent of which is used captively to produce unsaturated polyester resins, alkyd resins, and plasticizers.[5]

- Tenneco Chemical, Inc.

  Less than 20 percent of their 12-Gg/yr (2.6×10^7 lb/yr) MA production is used captively to produce fumaric acid, dibutyl maleate, and dodecanylsuccinic anhydride.[5]

- United States Steel Corporation

  Their MA capacity was recently expanded to 36 Gg/yr (7.9×10^7 lb/yr). Approximately 20 percent of their MA production is used captively to produce fumaric acid, dibutyl maleate, and dioctyl maleate.[5]

The expansion capabilities of 14 Gg (3.1×10^7 lb) each for Amoco and Ashland represent a current potential nationwide capacity of 267 Gg (58.9×10^7 lb).[2]

5.2.2 MA Usage and Demand

Maleic anhydride is an important raw material in the production of polyester resins, agricultural chemicals, lubricants, fumaric acid, copolymers, and other intermediate raw materials. Table 5-2 breaks out these categories by percentage and growth rates. Demand for MA has increased historically at a rate of 9 percent a year and is expected to grow at a rate of 6 to 11 percent over the next 5 years.[1] A major reason for the excess capacity is investor confidence that MA demand will be strong.

The predominant end use of MA is the production of unsaturated polyester resins, which go into reinforced plastic applications such as marine

TABLE 5-2. MALEIC ANHYDRIDE USAGE AND GROWTH

End use	1978 demand Gg/yr (lb/yr)	1978 demand as % of Production	Average annual % growth 1978-83
Unsaturated polyester resins	77.7 (1.7×10^8)	58	13
Agricultural chemicals	13.4 (3.0×10^7)	10	10
Lubricating additives	8.7 (1.9×10^7)	7	12
Fumaric acid	6.7 (1.5×10^7)	5	5
Copolymers	5.9 (1.3×10^7)	4	9
Maleic acid	4.2 (9.3×10^6)	3	10
Reactive plasticizers	4.0 (8.8×10^6)	3	8
Surface-active agents	3.2 (7.0×10^6)	2	8
Alkyd resins	1.5 (2.6×10^6)	1	4
Chlorendic anhydride and acid	1.2 (2.6×10^6)	1	13
Other	7.5 (1.7×10^7)	6	5
All MA products Total	134.0 (3.0×10^8)	Total 100	Ave. 11

SOURCES: Chemical Profile on Maleic Anhydride. Chemical Marketing Reporter. February 18, 1978.

Blackford, J. C. Marketing Research Report on Maleic Anhydride. Chemical Economics Handbook. Stanford Research Inst. Menlo Park, Calif. July 1976.

craft, building panels, automobiles, tanks, and pipes. Of the 134 Gg (30.0×10^7 lb) of maleic anhydride produced in 1978, polyester resins consumed 58 percent (78 Gg [17.0×10^7 lb]). Industry forecasters expect polyester production to grow at an annual rate of 10 to 15 percent. The growth rate would double this industry's use of maleic anhydride by 1981 to about 156 Gg (34.4×10^7 lb)--or more, if potential polyester markets in housing, marine craft, and autos materialize. Automobile manufacturers, striving to lighten products in order to meet Congressionally mandated fuel economy standards, are turning more and more to polyester with above average maleic content to replace heavier metals.[1] Presently, no chemical can substitute for MA, as in the production of polyester resins.

The agricultural chemicals' market, which is the second largest MA market (10 percent of demand), is expected to grow at a rate almost as rapid as polyester resins. This growth could be further accelerated by MA's use as a feedstock for agricultural pesticides. Although other chemicals can substitute for MA as an agricultural chemical, MA is highly competitive in this market.[1]

Other markets, such as lubricants, maleic anhydride copolymers, fumaric acid, and reactive plasticizers are expected to grow at either modest or rapid rates in the next 5 years.

In periods of excess capacity, MA is not considered a regional product. More than 15 percent of U.S. markets lie in each of the following regions: Middle Eastern, South Eastern, Western South Central, and Eastern North Central.[7] Polyester resins are primarily produced in the Central States, while agricultural chemicals and fumaric acid are mostly produced in the East. Although these geographic tendencies exist, MA is a homogeneous product and can be sold in any of the markets mentioned above.

5.2.3 Prices

5.2.3.1 Price of MA. Historically, prices of maleic anhydride have fluctuated widely (Figure 5-3); it sold at 37.4¢/kg (17.0¢/lb) in 1971 and at 28¢/kg (12.79¢/lb) in 1972. In these years, there was a small difference between the price costed by the producer (list price) and the price received (actual price). But at the end of 1973, the MA price began to rise, in large part due to increased benzene prices. List prices climbed steadily until they leveled off at 81.4¢ (37.0¢/lb) per molten kilogram and

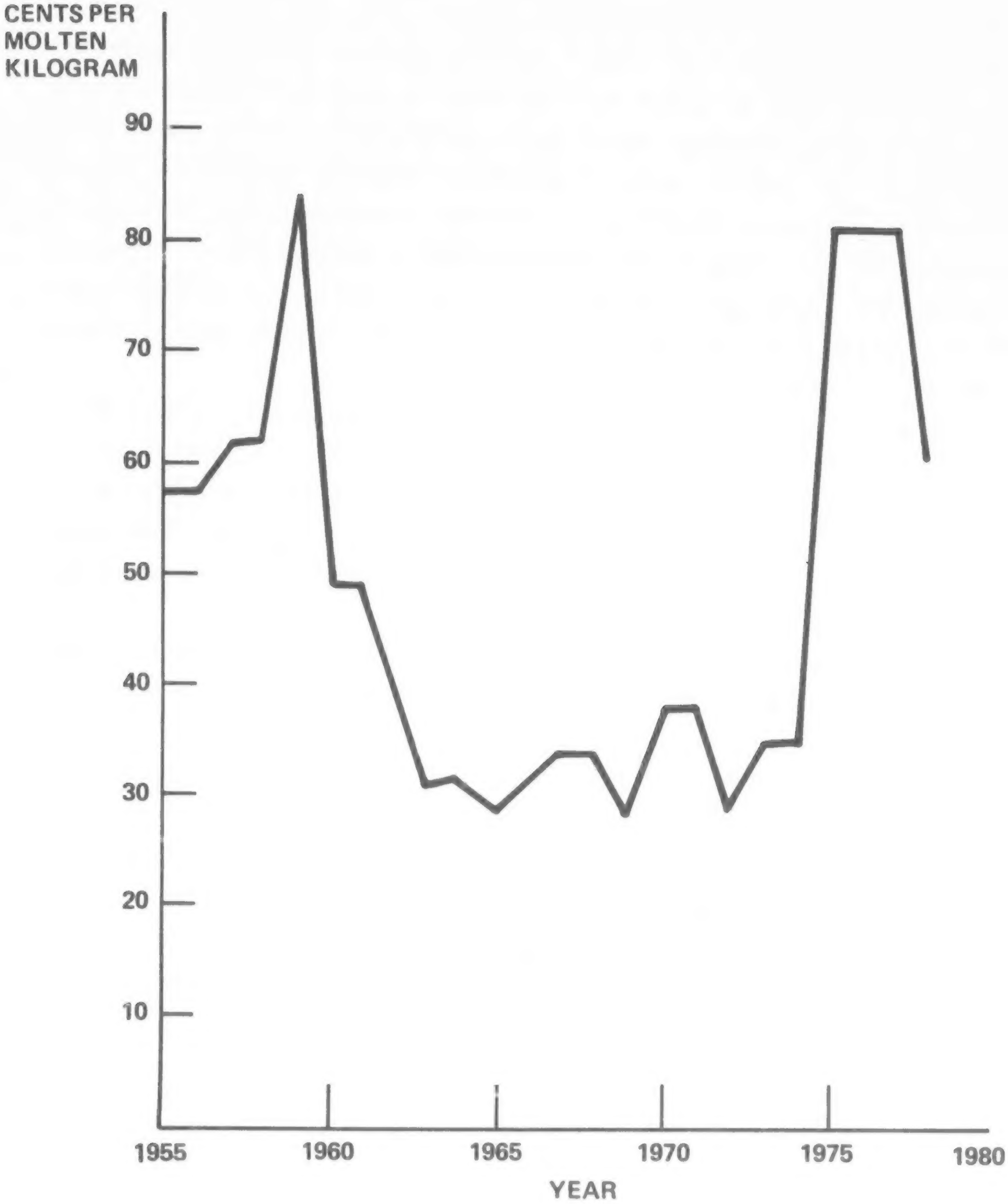

Figure 5-3. Price fluctuations of maleic anhydride.

SOURCE: Blackford, J. C. Marketing Research Report on Maleic Anhydride. Chemical Economics Handbook. Stanford Research Inst. Menlo Park. Calif. July 1976.

90.2¢ per bagged kilogram (41.0¢/lb).[1] Actual prices began to fall in 1977, leveling off at 61.7¢/kg (28.0¢/lb) per molten kilogram. This price held until the spring of 1979, at which time actual prices rose to 88¢/kg (40.0¢/lb).[8] In July 1979, MA prices rose once again to 96.8¢/kg (43.9¢/lb).[8] The ability of producers to raise prices this year has been attributed to increased MA demand and higher production costs, caused primarily by the soaring prices of benzene feedstock.

5.2.3.2 Feedstock Costs. Traditionally, costs for benzene feedstock accounted for 40 to 50 percent of the final price, but the quadrupling of world oil prices altered this relationship. Feedstock costs presently make up a 43- to 57-percent share of the production price, and this share will rise with increases in oil prices.[9] In 1978, the use of n-butane as a feedstock led to as much as a 7.3¢ reduction per kilogram of MA produced (3.3¢/lb) below benzene feedstock costs.[9] Since then, the price of benzene has increased at a faster rate than the price of n-butane. Thus, the cost differential between the two feedstocks has widened. If this trend continues, it may be economically favorable to use n-butane as the primary MA feedstock.

5.2.3.3 Transportation Costs. MA is usually transported by truck or by train. The price for transporting MA to its markets around the country has varied historically to within a few cents. It is several tenths of a cent per kilogram cheaper to transport maleic by train than by truck. However, a client must purchase at least 45,000 kg (1×10^5 lb) of maleic, the amount needed to fill a tank car, to make transportation by train practical. In 1978, costs for transporting MA by train averaged 0.005¢/kg/mi, whereas the costs for truck transport averaged 0.01¢/kg/mi.[10]

During periods of excess capacity, it is a common practice among producers to charge customers equal transportation costs regardless of location. Thus, the price paid by the customer will frequently not cover the actual cost of transportation. Domestic producers hope, however, that as MA market conditions improve, they can shift the cost of transportation from themselves to the customer.

5.2.4 Briquettes vs. Molten MA

The MA market can be segmented into two parts--briquette users and molten MA users. A typical briquette user is much smaller than a molten user and can neither buy in bulk (i.e., 45,000 kg [1×10^5 lb]) per tank car) nor afford liquid storage facilities, even though briquettes normally cost 6.5¢ to 9¢/kg (3.0¢ to 4.1¢/lb) more than the molten form and require up to 4.5¢/kg (2.0¢/lb) to melt down before use.[11] In addition, briquettes cannot be stored for periods longer than about 60 days because they begin to emit gases.[11]

Briquettes would be purchased by large users only in abnormal circumstances where foreign sources of briquettes temporarily undercut molten MA prices by more than the cost of melting down the MA.[11] This actually occurred recently and will be discussed in the following subsection on imports.

5.2.4.1 Imports Of Maleic Anhydride. Figure 5-4 shows the relationship between domestic production and demand and imports. All imported MA is in the form of briquettes since molten MA cannot be transported overseas. In 1978, imports accounted for less than 1 percent of the U.S. MA market but nearly 20 percent of the U.S. briquette market (see Figure 5-5).

In 1977, MA imports posed problems for U.S. producers. That year, imports contributed to the drop in price of domestic briquettes from 90¢ to 75¢/kg (40.9¢ to 34.1¢/lb) and the reduction of molten MA prices from 81.4¢ to 61.7¢/kg (36.9¢ to 28.0¢/lb). Although domestic overcapacity was undoubtedly the main factor for this price drop, briquettes were temporarily almost 4¢/kg (1.8¢/lb) cheaper to buy and melt down than the molten form.[11]

The low price of MA imported in 1977 was attributed to two possible factors:

- Foreign MA companies have excess capacity, so they purposely undercut U.S. prices in order to attract business; and
- Because many foreign companies may be partially subsidized by their governments, they can afford to sell at minimal or no operating profit.

In 1978, the quantity of imported MA declined considerably, from 3.5 Gg (7.6×10^6 lb) in 1977 to 1.0 Gg (2.3×10^6 lb) in 1978.[12] Several equally significant factors contributed to this reduction. First, benzene prices have risen abroad. Although benzene prices have increased domestically as well, foreign benzene prices have historically been and continue to be higher. Second, demand for MA has risen abroad. Consequently, foreign MA producers have been able to increase the prices they charge their custo-

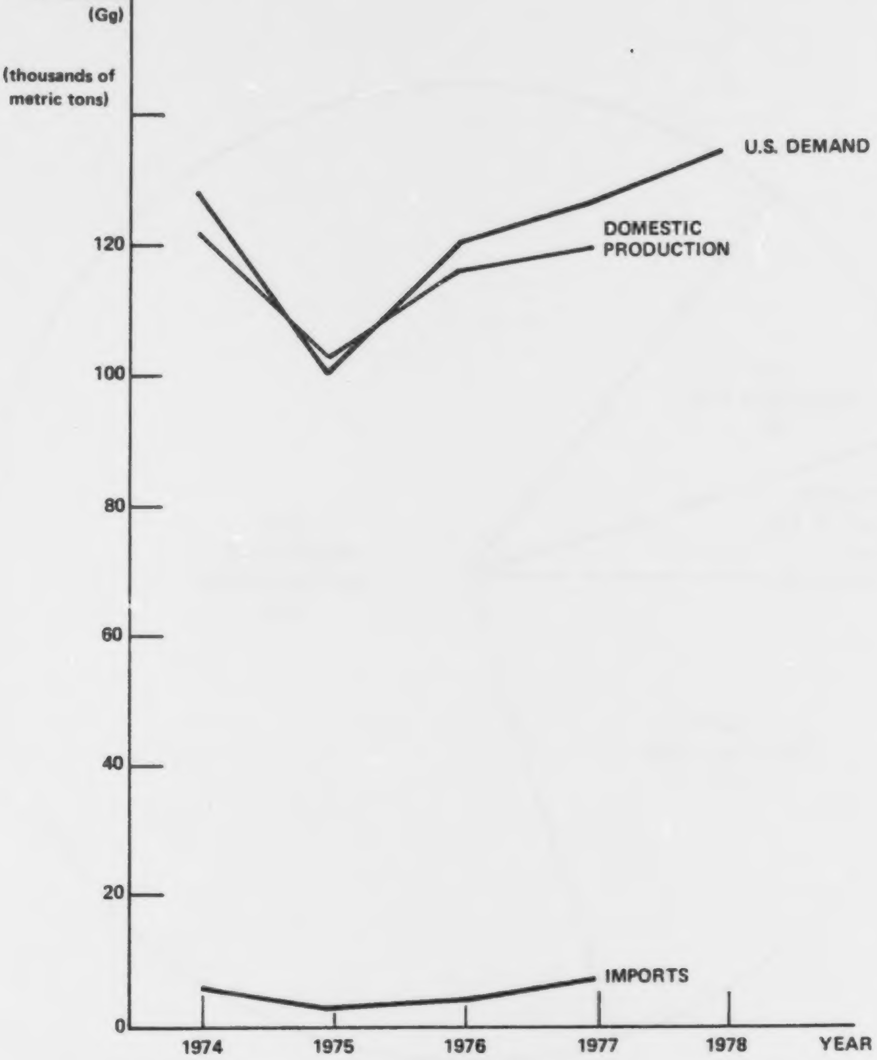

Figure 5-4. Maleic anhydride: a comparison of imports to U.S. production and demand.

SOURCE: Telecon. Epstein, E. A., Energy and Environmental Analysis, Inc., with Mr. Kendrik, DENKA. May 4, 1978.

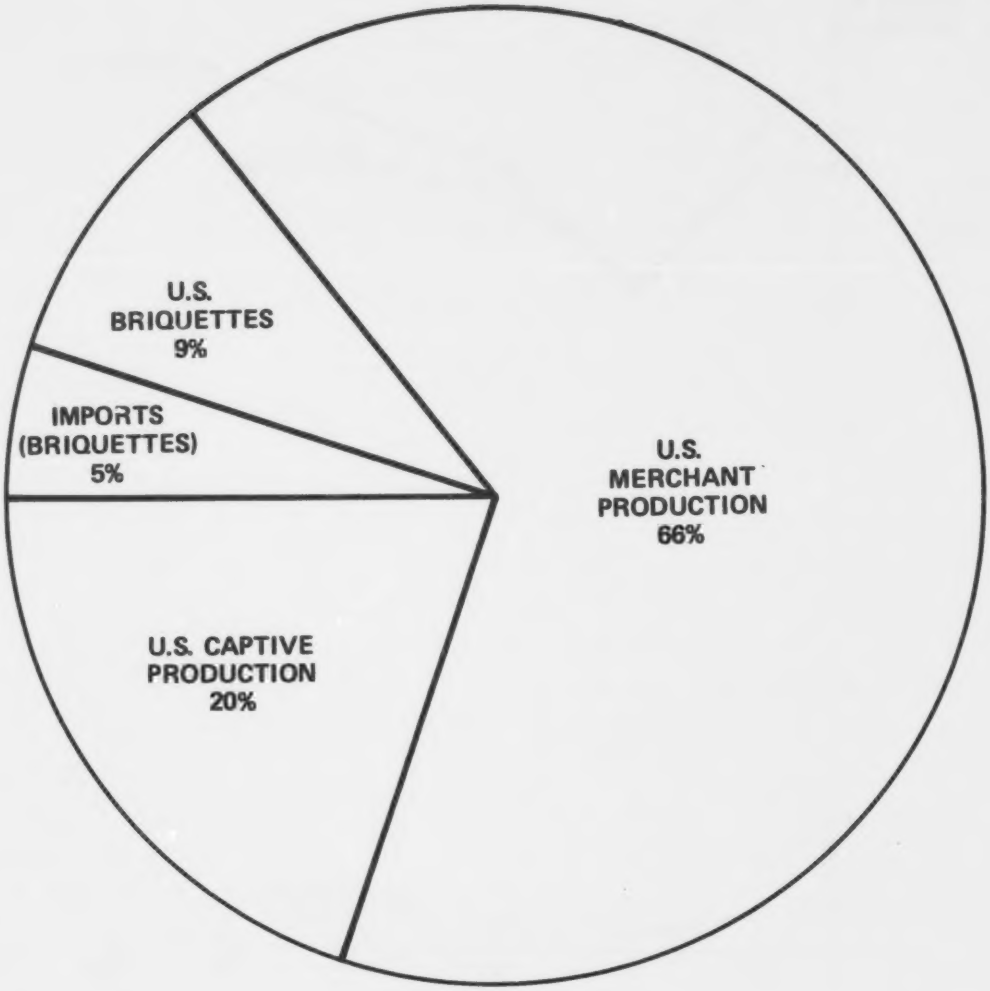

Figure 5-5. U.S. consumption of MA by source in 1977.

SOURCE: Telecon. Epstein, E.A., Energy and Environmental Analysis, Inc., with Magnusson, Fred, U.S. Department of Commerce. March 3, 1978.

mers as well as decrease their surplus supplies of MA. These price increases have brought foreign prices closer in line with domestic prices, thereby reducing the likelihood that foreign prices will continue to undercut the domestic market.

At present, equal quantities of butene (C_4)-based MA from Korea and Japan and benzene-based MA from Mexico and Italy are currently being imported. Of these, the C_4-based MA has the most favorable duty--6 percent ad valorem. Some domestic producers fear that this percentage is not high enough to keep large amounts of C_4-based MA from being imported in the future at considerably lower than current prices.

The import rate of duty on benzene-derived products is presently 3.8¢/kg (1.7¢/lb) plus 12.5 percent ad valorem. Prior to 1968, the rate was 7.7¢/kg (3.5¢/lb) plus 25 percent ad valorem; the duty, though lower, is still significant.[12]

Imports of benzenoid products from less developed countries (LDC's) are duty free, however, under the general system of import preferences. This means that imports from Mexico and Korea are not subject to duty as long as at least 35 percent of the value of their product comes from that country.[11] This raises the possibility that exporters might try to conceal the actual origin of some MA imports to obtain a low duty rate or avoid it altogether. The more important observation is that the two largest exporters of MA to the United States have no duty.

5.3 COST ANALYSIS OF ALTERNATIVE EMISSION CONTROL SYSTEMS

5.3.1 Introduction

This section presents costs for controlling and monitoring benzene emissions, where necessary, at each of the eight existing maleic anhydride plants using benzene as a feedstock. For each plant, two kinds of add-on control systems have been examined: thermal incineration with primary heat recovery and carbon adsorption. For each system, two benzene emission control levels--97 and 99 percent--have been costed. Monitoring costs have not been incorporated into the investment costs associated with each control system. They are discussed later in this section.

5.3.2 Summary of Technical Parameters Used as the Basis in Cost Analysis

This section summarizes the assumptions used in developing control costs at the 97- and 99-percent control level for eight maleic anhydride plants. It should be recognized, however, that these costs are only preliminary estimates (±30 percent). When these costs were developed, plant capacity was the only plant-specific parameter taken into consideration. To develop definite costs for an actual installation, a detailed engineering evaluation is required. Such an evaluation is beyond the resources and scope of this document.

The efficiencies and other parameters used in determining cost for the control methods in this analysis are listed in Table 5-3. Because some of these parameters vary from plant to plant, expressions have been derived in terms of plant capacity, "P," for the sake of brevity. Because benzene-fed maleic anhydride plants have similar process designs,[6] it has been assumed that the gas volumetric flow rate and the benzene emission rate vary in proportion to plant capacity.

Other assumptions used in determining cost for the add-on systems for each plant follow:

- Intensive stream parameters, such as gas pressure and temperature, are the same for all plants;
- No credit is given for control equipment used in an existing plant, unless the controls already achieve the alternative in question (i.e., 97 or 99 percent);
- All control system installations are retrofits, whose costs include retrofit penalties of 40 and 30 percent of the new plant installation cost for incinerator and carbon adsorption systems, respectively.[6] (These moderate penalties, in turn, reflect the fact that the costs of retrofitted systems are only somewhat greater than those for completely new plant installations. The primary retrofit difficulty may be finding adequate space to fit the control system into the existing plant layout.)

In addition to developing control costs for the various maleic anhydride plants, costs were developed for continuous monitoring of benzene stack emissions. The device costed is a gas chromatograph with appropriate auxiliary equipment including air sampler, data processor, and piping.

The add-on control costs have been based primarily on data available from an EPA contractor, Hydroscience, Inc.,[6] and a compendium of costs for selected air pollution control systems.[13] Monitoring costs were obtained from a vendor.[14]

TABLE 5-3. ESTIMATED TECHNICAL PARAMETERS USED IN DEVELOPING CONTROL SYSTEM COSTS

Parameter	Value[a]
1. Gas temperature	38° C (100° F)
2. Gas pressure	120 KPa (18 psia)
3. Gas volumetric flow rate	0.0536 P m^3/min (1.89 P ACFM)[b]
4. Inlet benzene emission rate	0.0084 P kg/hr (0.0185 P lb/hr)[b]
5. Benzene control efficiency	97%, 99%
6. Plant capacities	(See Tables 5-5 to 5-6)
7. Incinerator combustion temperature and residence time	760° C (1,400° F), 0.5 sec for 97% 870° C (1,600° F), 0.5 sec for 99%
8. Design carbon loading	6 kg VOC/100 kg carbon (6 lb VOC/ 100 lb carbon)

[a] EPA estimates.

[b] P = plant capacity in megagrams of maleic anhydride per year.

SOURCES: Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 28, 1978.

Two cost parameters--installed capital and total annualized--have been evaluated in this analysis. The installed capital cost for each emission control system includes the purchased costs of the major and auxiliary equipment, costs for site preparation and equipment installation, and design engineering costs. No attempt has been made to include costs for research and development, possible lost production during equipment installation, or losses during startup. Capital and operating costs in this section were developed in detail with cost parameters originally indexed to the fourth quarter of 1977. Because of rapid escalation of benzene and energy costs, the cost parameters were updated to the second quarter of 1979 without reoptimizing design parameters. Both the older and later data are shown in Table 5-4.

The total annualized cost (TAC) consists of direct operating costs, annualized capital charges, and recovery credits. Direct operating costs (DOC) include fixed and variable annual costs, such as:

- Labor and materials needed to operate control equipment;
- Maintenance labor and materials;
- Utilities, such as natural gas and electric power; and
- Liquid waste disposal.

The annualized capital charges account for depreciation, interest, administrative overhead, property taxes, and insurance. Depreciation and interest have been computed by use of a capital recovery factor, the value of which depends on the depreciable life of the control system and the annual interest rate (10 percent is used for the latter). Administrative overhead, taxes, and insurance have been fixed at an additional 4 percent of the installed capital cost.

The recovery credits apply to the value of material or energy recovered by the control system. With carbon adsorption systems, benzene is recovered from the regenerated carbon beds, while a credit for natural gas is applied to thermal incineration systems with primary heat recovery.

Finally, the total annualized cost is obtained by adding the direct operating costs to the annualized capital charges and subtracting the recovery credits from this sum.

TABLE 5-4. COST PARAMETERS[a]

Parameter	Value	
1. Operating factors	4,500 and 8,000 hr/yr	
2. Maintenance:		
Control systems	5.0% of total installed cost	
Monitoring system	3.4% of total installed cost	
3. Depreciation and interest	16.28%[b] of total installed cost	
4. Taxes, insurance, overheads	4.0% of total installed cost	
	Fourth quarter 1977	Second quarter 1979
5. Utilities:		
Electricity	\$0.03/kWh	\$0.04/kWh
Natural gas	\$1.90/GJ (\$2.00/MM Btu)	\$2.85/GJ (\$3.00/MM Btu)
Steam	\$5.50/Mg (\$2.50/M lb)	\$8.15/Mg (\$3.70/M lb)
Cooling water	\$0.026/kL (\$0.10/M gal)	\$0.037/kL (\$0.14/M gal)
6. Operating materials		
Sodium hydroxide (50%)	\$0.20/kg (\$0.09/lb)	\$0.20/kg (\$0.09/lb)
Activated carbon	\$1.90/kg (\$0.85/lb)	\$2.42/kg (\$1.10/lb)
7. Liquid waste disposal	\$2.60/kL (\$10/M gal)	\$2.60/kL (\$10/M gal)
8. Credits		
Primary heat recovery	Value of natural gas	
Recovered benzene	\$0.17/L (\$0.63/gal)	\$0.33/L (\$1.25/gal)
9. Plant cost index	210	237

[a] EPA estimates.

[b] Based on a 10-year life and a 10-percent annual interest.

NOTE: Although a plant does not shut down for 3,500 hr/yr, when demand is reduced, it is easier to estimate costs this way. This is not expected to change the estimated costs appreciably.

SOURCES: Benzene Price Quotations. Chemical Marketing Reporter. June 18, 1979.

Economic Indicators. Chemical Engineering. 86(13):7. 1979.

Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Joe Lorber, Hewlett-Packard. March 22, 1978.

5.3.3 Control Costs for Maleic Anhydride Facilities

Costs for controlling benzene emissions have been developed for each of the eight benzene-fed maleic anhydride plants for the two regulatory options of 97 and 99 percent control. Only existing plant costs are included because no new plants are anticipated before 1983. The maleic anhydride industry has been operating at 56 percent capacity. Projected increases in demand could therefore be accommodated by increases in operating rates over the next few years.

5.3.3.1 Costs to Achieve the 97-Percent Regulatory Option. The costs for reducing benzene emissions by 97 percent are shown in Tables 5-5a and 5-5b for the carbon adsorption and thermal incineration control systems, respectively.

Of the eight plants for which costs are shown, the best information available indicates that three (DENKA, Reichhold in Elizabeth, N.J., and Koppers) already reduce their benzene emissions by at least 97 percent. Therefore, the control costs are indicated for only the five other plants. Of the two control systems costed at 97 percent, the carbon adsorption systems are inherently more complex and more expensive. But the adsorption systems have lower direct operating costs, which are, at mid-1979 prices for benzene, tempered by credits for recovered benzene. As a result, the total annualized costs are somewhat lower for the adsorption systems at full capacity and nearly equal at 4,500 hr/yr.

5.3.3.2 Costs to Achieve the 99-Percent Regulatory Option. Tables 5-6a and 5-6b display respective control costs to achieve 99 percent benzene emission reduction via carbon adsorption and thermal incineration, with heat recovery. Except for the Koppers plant, all benzene-fed plants require additional control to achieve this level. Table 5-6b lists costs for a thermal incinerator with primary heat recovery. As capacity increases from 11,800 to 38,500 Mg/yr (26 to 85 million lb/yr), the installed cost increases from $0.78 million to $1.88 million--only slightly more than the incineration capital costs in Table 5-5b, which were for 97 percent control. However, because of high fuel costs, the direct operating costs for incineration at 99 percent control are significantly higher than those for 97 percent control. These costs range from about $0.68 to $2.15 million/yr at 8,000 hr/yr.

TABLE 5-5a. EXISTING PLANT COSTS FOR ACHIEVING 97 PERCENT BENZENE EMISSION REDUCTION CONTROL METHOD: CARBON ADSORPTION

				Direct operating cost ($/yr)[a]		Benzene recovery credit ($/yr)[a]		Total annualized cost[b] ($/yr)[a]		Total annualized cost[b] ($/Mg product)	
Plant name and location	Capacity (Mg/yr)	Total installed cost ($)[a,c]	ACC ($/yr)[a]	4,500 hr/yr[d]	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr
1. Ashland--Neal, W. Va.	27,200	1,650	335	467	739	(374)	(664)	428	410	28.6	15.4
2. Monsanto--St. Louis, Mo.	38,100[e]	2,240	454	637	1,018	(523)	(930)	568	542	26.9	14.5
3. DENKA--Houston, Tex.[f]	22,700	0	0	0	0	0	0	0	0	0	0
4. Reichhold--Elizabeth, N.J.[f]	13,600	0	0	0	0	0	0	0	0	0	0
5. Reichhold--Morris, Ill.	20,000	1,250	254	354	554	(274)	(488)	334	320	30.5	16.4
6. Tenneco--Fords, N.J.	11,800	780	158	225	334	(162)	(288)	221	213	34.7	18.8
7. U.S. Steel--Neville Island, Pa.	38,500	2,260	458	643	1,028	(528)	(939)	573	547	26.9	14.5
8. Koppers--Bridgeville, Pa.[f]	15,400	0	0	0	0	0	0	0	0	0	0

[a] Dollars are stated in thousands.

[b] Includes the cost of benzene continuous monitoring (approximately $9,000/yr per plant).

[c] Installed costs are rounded to the nearest 10 thousand dollars; other costs are rounded to the nearest thousand.

[d] Operating factor represents production at 56 percent of capacity.

[e] This represents 80 percent of the total plant capacity. The rest of the MA is produced from n-butane.

[f] These plants already achieve 97 percent reduction.

SOURCES: Benzene Price Quotation. Chemical Marketing Reporter. June 18, 1979.

Blacker, Herbert G., and Thomas M. Nichols. Capital and Operating Costs of Pollution Control Equipment Modules--Vol. II--Data Manual. ICARUS Corp. Silver Spring, Md. (Prepared for Office of Research and Monitoring, U.S. Environmental Protection Agency. Washington, D.C.) EPA-R5-73-023b. July 1973.

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Kinkley, M. T., and R. B. Neverill. Capital and Operating Costs of Selected Air Pollution Control System. GARD, Inc. Niles, Ill. EPA-450/3-76-014. May 1976.

Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 28, 1978.

Letter from Vatavuk, William M., to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. June 27, 1978.

TABLE 5-5b. EXISTING PLANT COSTS FOR ACHIEVING 97 PERCENT BENZENE EMISSION REDUCTION CONTROL METHOD: THERMAL INCINERATION WITH PRIMARY HEAT RECOVERY

Plant name and location	Capacity (Mg/yr)	Total installed cost ($)[a,c]	ACC ($/yr)[a]	Direct operating cost ($/yr)[a]		Benzene recovery credit ($/yr)[a]		Total annualized cost ($/yr)[a]		Total annualized cost ($/Mg product)[b]	
				4,500 hr/yr[d]	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr
1. Ashland--Neal, W. Va.	27,200	1,320	260	701	1,176	(528)	(938)	433	498	28.9	18.6
2. Monsanto--St. Louis, Mo.	38,100[e]	1,730	351	966	1,632	(739)	(1,314)	578	669	27.3	17.8
3. DENKA--Houston, Tex.[f]	22,700	0	0	0	0	0	0	0	0	0	0
4. Reichhold--Elizabeth, N.J.[f]	13,600	0	0	0	0	0	0	0	0	0	0
5. Reichhold--Morris, Ill.	20,000	1,070	216	503	876	(388)	(690)	331	402	30.2	20.6
6. Tenneco--Fords, N.J.	11,800	730	149	302	531	(229)	(407)	222	273	34.8	23.9
7. U.S. Steel--Neville Island, Pa.	38,500	1,750	355	976	1,650	(747)	(1,328)	584	677	27.3	17.8
8. Koppers--Bridgeville, Pa.[f]	15,400	0	0	0	0	0	0	0	0	0	0

[a] Dollars are stated in thousands.

[b] Includes the cost of benzene continuous monitoring (approximately $9,000/yr per plant).

[c] Installed costs are rounded to the nearest 10 thousand dollars; other costs are rounded to the nearest thousand.

[d] Operating factor represents production at 56 percent of capacity.

[e] This represents 80 percent of the total plant capacity. The rest of the MA is produced from n-butane.

[f] This plant already achieves 97 percent reduction.

SOURCES: Economic Indicators. Chemical Engineering. 86(13):7. 1979.

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Letter from Hewett, P. S., Reichhold Chemicals, Inc., to Patrick, D. R., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 27, 1978.

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Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 28, 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. June 27, 1978.

TABLE 5-6a. EXISTING PLANT COSTS FOR ACHIEVING 99 PERCENT BENZENE EMISSION REDUCTION CONTROL METHOD: CARBON ADSORPTION

Plant name and location	Capacity (Mg/yr)	Total installed cost ($)[a,c]	ACC ($/yr)[a]	Direct operating cost ($/yr)[a]		Benzene recovery credit ($/yr)[a]		Total annualized cost ($/yr)[a]		Total annualized cost ($/Mg product)[b]	
				4,500 hr/yr[d]	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr
1. Ashland--Neal, W. Va.	27,200	1,680	341	481	764	(382)	(680)	440	425	29.3	16.0
2. Monsanto--St. Louis, Mo.	38,100[e]	2,270	460	657	1,052	(536)	(952)	581	560	27.5	14.9
3. DENKA--Houston, Tex.	22,700	1,440	292	409	645	(319)	(568)	382	369	30.6	16.7
4. Reichhold--Elizabeth, N.J.	13,600	900	183	261	402	(191)	(340)	253	245	34.2	18.7
5. Reichhold--Morris, Ill.	20,000	1,280	260	365	573	(281)	(500)	344	333	31.4	17.1
6. Tenneco--Fords, N.J.	11,800	810	164	232	355	(166)	(295)	230	224	36.0	19.7
7. U.S. Steel--Neville Island, Pa.	38,500	2,290	464	663	1,063	(541)	(962)	586	565	27.5	14.9
8. Koppers--Bridgeville, Pa.[f]	15,400	0	0	0	0	0	0	0	0	0	0

[a] Dollars are calculated by thousands.

[b] Includes the cost of benzene continuous monitoring (approximately $9,000/yr per plant).

[c] Installed costs are rounded to the nearest 10 thousand dollars; other costs are rounded to the nearest thousand.

[d] Operating factor represents production at 56 percent of capacity.

[d] This represents 80 percent of the total plant capacity. The rest of the MA is produced from n-butane.

[f] This plant already achieves 99 percent reduction.

SOURCES: Benzene Price Quotation. Chemical Marketing Reporter. June 18, 1979.

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Kinkley, M. T., and R. B. Neverill. Capital and Operating Costs of Selected Air Pollution Control System. GARD, Inc. Niles, Ill. EPA-450/3-76-014. May 1976.

Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

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Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, March 8, 1978.

TABLE 5-6b. EXISTING PLANT COSTS FOR ACHIEVING 99 PERCENT BENZENE EMISSION REDUCTION CONTROL METHOD: THERMAL INCINERATION WITH PRIMARY HEAT RECOVERY

				Direct operating cost ($/yr)[a]		Energy recovery credit ($/yr)[a]		Total annualized cost ($/yr)[a]		Total annualized cost ($/Mg product)[b]	
Plant name and location	Capacity (Mg/yr)	Total installed cost ($)[a,c]	ACC ($/yr)[a]	4,500 hr/yr[d]	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr	4,500 hr/yr	8,000 hr/yr
1. Ashland--Neal, W. Va.	27,200	1,400	284	899	1,526	(625)	(1,111)	558	699	37.1	26.0
2. Monsanto--St. Louis, Mo.	38,100[e]	1,860	378	1,246	2,124	(876)	(1,557)	748	945	35.3	25.0
3. DENKA--Houston, Tex.	22,700	1,220	247	757	1,280	(522)	(927)	482	600	38.5	26.8
4. Reichhold--Elizabeth, N.J.	13,600	850	173	468	782	(313)	(557)	328	398	44.1	29.9
5. Reichhold--Morris, Ill.	20,000	1,130	229	672	1,133	(460)	(817)	441	545	40.0	27.7
6. Tenneco--Fords, N.J.	11,800	780	158	411	683	(271)	(482)	298	359	46.3	30.4
7. U.S. Steel--Neville Island, Pa.	38,500	1,880	382	1,259	2,146	(885)	(1,573)	756	955	35.3	25.0
8. Koppers--Bridgeville, Pa.[f]	15,400	0	0	0	0	0	0	0	0	0	0

[a] Dollars are calculated by thousands.

[b] Includes the cost of benzene continuous monitoring (approximately $9,000/yr per plant).

[c] Installed costs are rounded to the nearest 10 thousand dollars; other costs are rounded to the nearest thousand.

[d] Operating factor represents production at 56 percent of capacity.

[e] This represents 80 percent of the total plant capacity. The rest of the MA is produced from n-butane.

[f] This plant already achieves 99 percent reduction.

SOURCES: Economic Indicators. Chemical Engineering. 86(13):7. 1979.

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Kinkley, M. T., and R. B. Neverill. Capital and Operating Costs of Selected Air Pollution Control System. GARD, Inc. Niles, Ill. EPA-450/3-76-014. May 1976.

Lawson, J. F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 28, 1978.

Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. March 8, 1978.

As in the incineration systems for 97 percent control, a heat exchanger is built into the incinerator unit to preheat the incoming air, while cooling the exhaust gas. In this way, about 50 percent of the total heat duty of the incinerator is recovered, as shown in Table 5-6b. Depending on plant size and operating mode, these credits amount to 70 to 80 percent of the direct operating costs.

The carbon adsorption system costs shown in Table 5-6a also represent achievement of a 99-percent control efficiency. Ranging from $0.81 to $2.29 million, the installed costs for the 99-percent adsorbers are moderately higher than those for the incineration systems. However, lower direct operating costs are generated for the adsorption systems for incineration, and a larger fraction (80 to 90 percent) of these operating costs is offset by credits for recovered benzene at today's prices. The end result is that the total annualized cost, for 99 percent control at full capacity, is conspicuously less for adsorption than for incineration: a range of $224,000 to $565,000/yr as opposed to a range of $359,000 to $955,000/yr. As production is cut back, the advantage of adsorption is gradually lost because the process is capital intensive. Yet even at 4,500 hr/yr, adsorption is preferable.

5.3.3.3 Comparison of Control Levels. When an incinerator design is revised from 760° C (1,400° F) for 97 percent control to 870° C (1,600° F) for 99 percent control, construction materials, insulation thickness, and firebox volume undergo changes that affect capital costs. Fuel expense and maintenance necessarily increase. The addition of a heat exchanger for nominal, 50-percent primary recovery incurs more capital expense, but this expense should be more than offset by fuel savings. (Strictly speaking, the optimal degree of primary heat recovery depends on scale, temperature, and anticipated fuel costs. No attempt has been made here to tailor recovery to the different situations.) The costs shown in Tables 5-5b (97 percent) and 5-6b (99 percent) reflect engineering estimates of these factors.

Unlike an incinerator, a carbon adsorption unit is not normally designed to a specified efficiency.[13] The underlying science and engineering are inexact, and the adsorptive capacity of the carbon diminishes unpredictably with use. If an adsorption system is built to operate as efficiently as possible, the net annualized operating costs will be insensitive to control level. It is almost certain, however, that the efficiencies required here are greater than the economic optimum and that 99 percent control costs more than 97 percent control. Certainly, design and operating changes exist that would increase both control efficiency and cost. To represent this effect meaningfully, in the absence of a strict design method, the amount of carbon required is assumed to be directly proportional to the desired control efficiency. (Clearly, this assumption fails at 100 percent.) The capital and operating cost differentials between Tables 5-5a and 5-6a result from this assumption, but the end result is an annualized cost that is almost unaffected by the control level.

5.3.3.4 Monitoring. Table 5-7 contains costs for continuous monitoring of benzene stack emissions from both the 97- and 99-percent emission controls using a gas chromatograph with a flame ionization detector. The installed cost of the chromatograph and its auxiliaries is $35,000. Depending on the operating factor, the direct operating cost varies from about $1,450 to $1,680/yr. When the annualized capital charges are added, the total annualized cost amounts to $8,600 and $8,800/yr for the 4,500- and 8,000-hr/yr cases, respectively. These costs are relatively low when compared to the control costs in Tables 5-5 and 5-6, where the annualized monitoring cost has been rounded to $9,000 for all cases.

5.3.4 Cost Effectiveness of the Alternative Emissions Limits

For each of the control methods costed to achieve one or more of the alternative emission limits considered, it is informative to compare the total annualized cost with the amount of benzene removed. A convenient yardstick for expressing this comparison is the cost-effectiveness ratio, the quotient of the annualized cost and the quantity of benzene removed annually, expressed in dollars per megagram of benzene removed.

These ratios and other important cost data appear in the cost summary table, Table 5-8. Three kinds of data are shown: total annualized cost, the amount of benzene removed annually by the control method, and the cost-effectiveness ratio. Because the total annualized cost is expressed in dollars per megagram of maleic anhydride produced annually, two costs are given: one computed at 4,500 hr/yr operating factor, the other at full capacity, or 8,000 hr/yr.

TABLE 5-7. COSTS FOR CONTINUOUS MONITORING OF BENZENE STACK EMISSIONS[a]

Cost	Operating factor (hr/yr)	
	4,500	8,000
Installed ($)	35,000	35,000
Direct operating ($/yr)	1,450	1,680
Annualized capital ($/yr)	7,100	7,100
Total annualized ($/yr) (rounded)[b]	8,600	8,800

[a] These costs apply to all existing plants using benzene feed.

[b] Includes: gas chromatograph with flame ionization detector, automatic gas sampling value, air sampler, post-run calculator, and gas regulators.

SOURCE: Letter from Vatavuk, William M., Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, to Lorber, Joe, Hewlett-Packard. March 22, 1978. (Updated to mid-1979 by rules of thumb.)

TABLE 5-8. COST SUMMARY FOR EXISTING MALEIC ANHYDRIDE PLANTS

	97 Percent Benzene Emission Reduction											
	Thermal incineration with primary heat recovery						Carbon adsorption					
	4,500 hr/yr			8,000 hr/yr			4,500 hr/yr			8,000 hr/yr		
Plant name and location	TAC ($/Mg product)[a]	Benzene removed (Mg/yr)[b]	Cost-effectiveness ($/Mg)[c]	TAC	Benzene removed	Cost-effectiveness	TAC	Benzene removed	Cost-effectiveness	TAC	Benzene removed	Cost-effectiveness
Ashland	28.9	998	443	18.6	1,773	286	28.6	998	438	15.4	1,773	236
Monsanto (St. Louis, Mo.)	27.3	1,398	420	17.8	2,484	273	26.9	1,398	411	14.5	2,484	220
DENKA (Houston, Texas)	-	-	-	-	-	-	-	-	-	-	-	-
Reichhold (Elizabeth, N.J.)	-	-	-	-	-	-	-	-	-	-	-	-
Reichhold (Morris, Ill.)	30.2	734	463	20.6	1,304	314	30.5	734	467	16.4	1,304	292
Tenneco (Fords, N.J.)	34.8	433	533	23.9	769	367	34.7	433	531	18.8	769	290
U.S. Steel (Neville Island, Pa.)	27.3	1,413	420	17.8	2,510	273	26.7	1,413	411	14.5	2,510	220
Koppers (Pa.)[d]	-	-	-	-	-	-	-	-	-	-	-	-

	99 Percent Benzene Emission Reduction											
	Thermal incineration with primary heat recovery						Carbon adsorption					
	4,500 hr/ yr			8,000 hr/yr			4,500 hr/yr			8,000 hr/yr		
Plant name and location	TAC ($/Mg product)[a]	Benzene removed (Mg/yr)[b]	Cost-effectiveness ($/Mg)[c]	TAC	Benzene removed	Cost-effectiveness	TAC	Benzene removed	Cost-effectiveness	TAC	Benzene removed	Cost-effectiveness
Ashland (Neal, W. Va.)	37.1	1,018	557	26.0	1,810	391	29.3	1,018	441	16.0	1,810	240
Monsanto (St. Louis, Mo.)	35.3	1,397	542	25.0	2,484	384	27.5	1,397	422	14.9	2,484	229
DENKA (Houston, Texas)	38.5	832	590	26.8	1,480	411	30.6	832	470	16.7	1,480	255
Reichhold (Elizabeth, N.J.)	44.1	499	675	29.9	887	454	34.2	499	525	18.7	887	286
Reichhold (Morris, Ill.)	40.0	773	614	27.7	1,304	425	31.4	773	482	17.1	1,304	262
Tenneco (Fords, N.J.)	46.3	433	709	30.4	764	479	36.0	433	552	19.7	769	303
U.S. Steel (Neville Island, Pa.)	35.3	1,412	542	25.0	2,510	384	27.5	1,412	422	14.9	2,510	229
Koppers (Pa.)[d]	-	-	-	-	-	-	-	-	-	-	-	-

[a] Includes costs for continuous monitoring.

[b] Product of the inlet emission rate and the control efficiency.

[c] Quotient of the total annualized cost ($/yr) and the benzene removed (Mg/yr).

[d] Koppers is expected to incur no costs under either control level.

SOURCES: Blacker, Herbert G., and Thomas M. Nichols. Capital and Operating Costs of Pollution Control Equipment Modules--Vol. II--Data Manual. ICARUS Corp. Silver Spring, Md. (Prepared for Office of Research and Monitoring, U.S. Environmental Protection Agency. Washington, D.C.) Report Number EPA-R5-73-023b. July 1973.

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Kinkley, M. T., and R. B. Neverill. Capital and Operating Costs of Selected Air Pollution Control System. Hydroscience, Inc. EPA-450/3-76-014. May 1976.

Lawson, J. F. Emissions Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride--Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality, Planning and Standards, U.S. Environmental Agency. Research Triangle Park, N.C. EPA Contract Number 68-02-2577. March 1978.

Letter from Vatavuk, William M., Office of Air Quality, Planning, and Standards, U.S. Environmental Protection Agency, to Lorber, Joe, Hewlett-Packard. March 22, 1978.

Letter from Vatavuk, William M., Office of Air Quality, Planning, and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality, Planning, and Standards, U.S. Environmental Protection Agency. March 28, 1978.

Letter from Vatavuk, William M., Office of Air Quality, Planning, and Standards, U.S. Environmental Protection Agency, to Weber, Robert, Office of Air Quality, Planning, and Standards, U.S. Environmental Protection Agency. June 27, 1978.

Each "benzene-removed" number is the product of the inlet emission loading to the control device (see Table 5-3) and the control efficiency.

Figures 5-6 and 5-7 depict the cost-effectiveness ratios for the 4,500- and 8,000-hr/yr operating factors, respectively. To display the data better, the vertical axes on these figures have been expanded.

In Figures 5-6 and 5-7, note that the 99-percent curves lie above the 97-percent curves for incineration but not for adsorption. These results can be explained by considering the two control techniques employed--thermal incineration and carbon adsorption.

For incineration, the lower combustion temperature employed at the 97-percent control level (as opposed to the 99-percent level) leads to conspicuously lower capital and operating costs for the system, with only slight changes in the amount removed. Consequently, cost effectiveness improves somewhat with adoption of the 97-percent control level, although it might decrease at lower control levels.

For carbon adsorption, the similarity between the 97- and 99-percent control level costs is due to the assumptions described earlier, employed in estimating costs at 97 percent control. Neither capital nor operating costs, modified by benzene credits, change significantly between the two cases. This would not be true if one sought more stringent control than 99 percent.

Finally, note that in both Figures 5-6 and 5-7 the curves slope gradually downward with increasing plant capacity, indicating the expected economy of scale. For plant sizes above 40,000 Mg/yr (8.8 × 10^7 lb/yr), the cost-effectiveness ratios may seem to approach asymptotic values. If design parameters are held constant as plant capacity increases, the influence of the fixed costs on the total annualized cost becomes less and less pronounced, while the variable costs--those nearly proportional to production capacity--become more influential. In other words, the total annualized cost becomes nearly proportional to the production capacity. Then, because the amount of benzene removed is exactly proportional to production capacity, the ratio of these quantities approaches a constant value.

It is normal in process design, however, that such parameters as the liquid/gas ratio in the reactor vent scrubber or the fractional heat recovery in the incinerator, assume different optimal values at different scales.

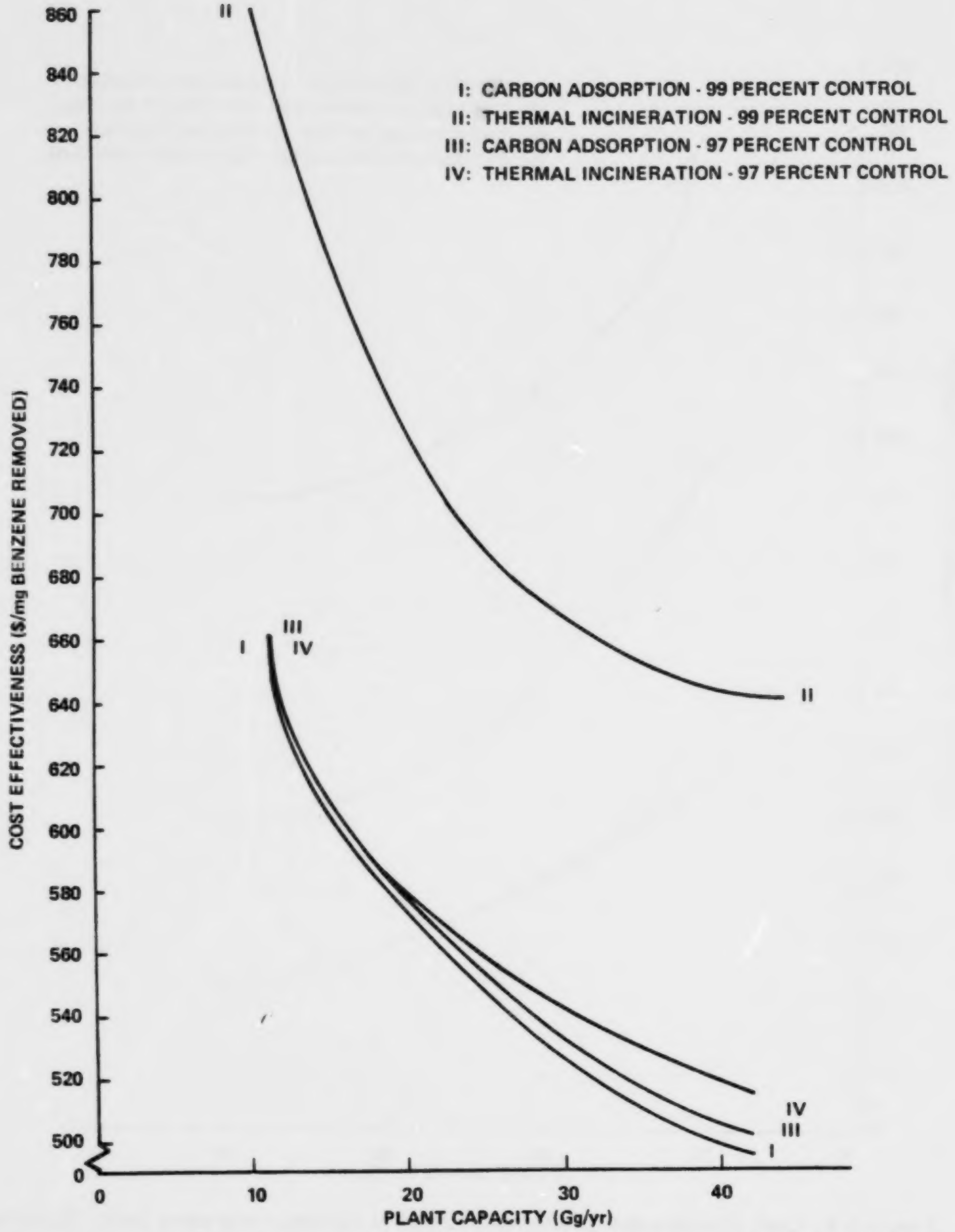

Figure 5-6. Cost effectiveness of alternative control systems—operating factor 4,500 hours, with monitoring.

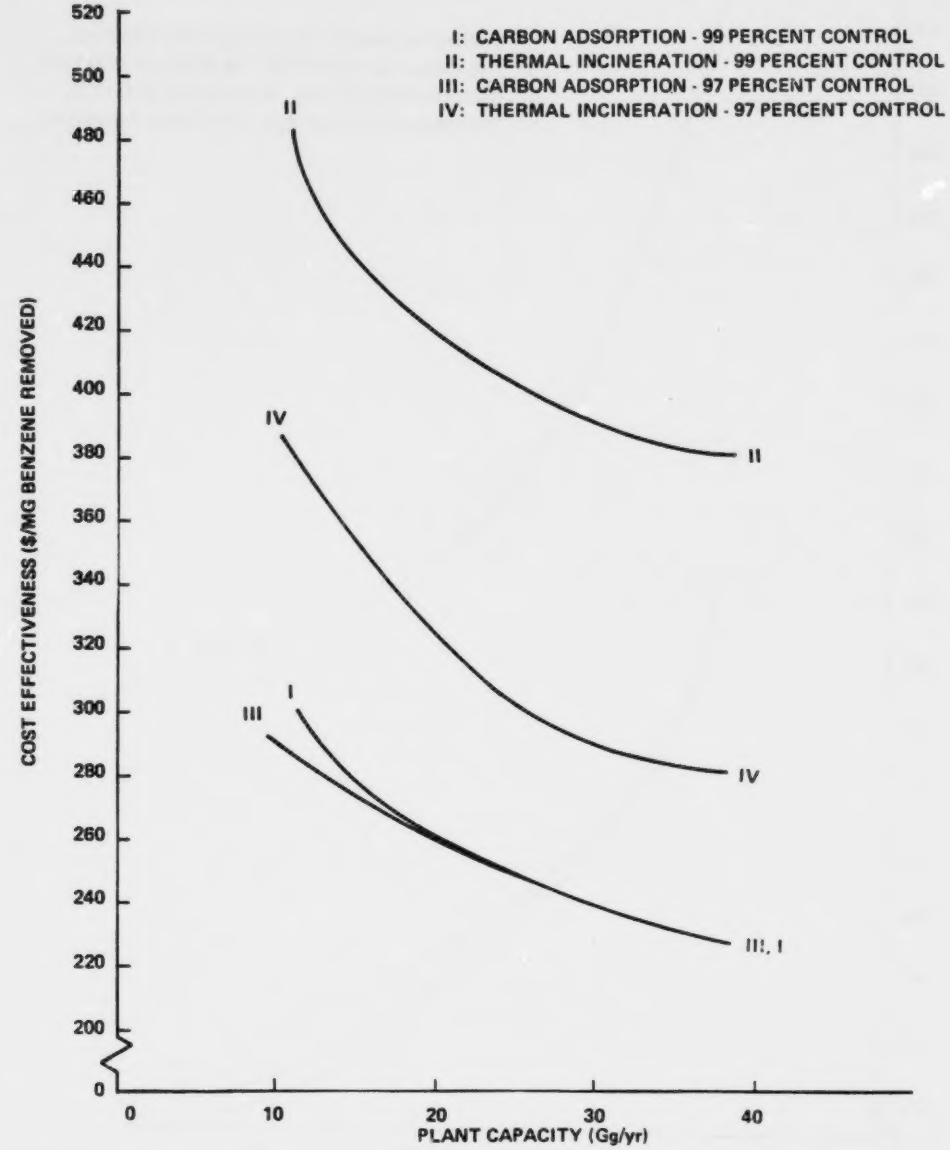

Figure 5-7. Cost effectiveness of alternative control systems—operating factor 8,000 hours, with monitoring.

Ordinarily, the consequence is that the large-scale version is more "efficient" in a variety of ways, and both fixed and variable costs increase less rapidly than productive capacity.

5.3.5 Control Cost Comparison

Before the accuracy and representativeness of control costs can be ascertained, they must be compared with costs obtained from other data sources. In doing this, one can either compare the installed capital costs, the annualized costs, or both. However, because the capital costs influence the annualized costs (via the annualized capital charges) and because the other terms in the annualized cost contain more site-specific variability (utilities, for instance), it is preferable to limit the comparison to the installed costs.

Even for a control system sized for a particular emission point, the installed cost may vary considerably from site to site. Such factors as the cost of installation labor (electricians, pipefitters, etc.), the requirement of special installation materials (e.g., extra insulation for systems installed in colder climates), and the presence or absence of excess utility capacity considerably influence the total installed cost.

With this in mind, however, capital cost comparisons can be made among a range of control system sizes. Such comparisons are best made graphically; that is, installed costs adjusted to the same reference date (second quarter 1979, in this case) are plotted against some technical parameter relevant to the control system. In this section, installed costs are compared among various sizes of carbon adsorbers and thermal incinerators with primary heat recovery. The technical parameters used in the comparisons are the gas volumetric flow rates at the control system inlet and outlet, respectively.

For the carbon adsorbers, capital costs developed for the existing plants (Tables 5-5a and 5-6a) are compared with costs developed by an EPA contractor and with an actual cost for a system installed at a maleic anhydride plant. These costs have been plotted against the inlet volumetric flow rate on logarithmic scales (Figure 5-8). Note that two EPA curves appear in the figure. The top EPA curve applies to a 99-percent efficient adsorber, while the one beneath it is a unit designed for 97 percent efficiency.

Consider first the 99-percent adsorbers. Figure 5-8 shows that the contractor's costs,[6] after adjustment to the same time base, exceed the EPA figures over the entire range of 730 to 2,000 m^3/min. However, the differences between the costs range only from 5 to 21 percent, which falls within the ±30 percent accuracy range assigned to the costs.

Figure 5-9 displays the installed costs for thermal incineration systems with primary heat recovery. The EPA cost at this flow rate, 5,920 m^3/min ($2.1 \times 10^5 ft^3$/min), is $1.40 million, which is $850,000 (60 percent) less than the industry figure. This discrepancy results primarily from a design difference between the two units. The plant system contains a boiler for heat recovery, while the EPA system uses a simple gas-to-gas heat exchanger. As expected, the former system is more costly.

On the other hand, EPA costs are significantly higher than a contractor's trend line shown in Figure 5-9, both of which represent 99 percent control efficiency. The cost difference varies from 74 to 76 percent over the flow rate range of 2,200 to 8,000 m^3/min (78,000 to 280,000 ft^3/min). After an examination of itemized costs from these two data sources, it appears that most of the discrepancy is caused by the costs of the respective incinerator chambers. In addition, the EPA figures include costs for ductwork and stacks, while the contractor costs do not. Differences arising from variation in scope are not unusual.

5.4 ECONOMIC IMPACT OF REGULATORY OPTIONS

5.4.1 Introduction

This section discusses the economic impact of two possible regulatory options for existing MA plants--97 percent and 99 percent benzene emission reductions. Subsection 5.4.2 describes the impacts on individual MA manufacturers because of increased capital budget requirements, shifts in competitive position due to unequal control costs, and other possible costs borne by MA producers to meet other environmental standards. Subsection 5.4.3 describes the impact on the price of products that use MA. Subsection 5.4.4 describes effects on employment and balance of trade. Finally, Subsection 5.4.5 summarizes the annualized costs incurred 5 years from now.

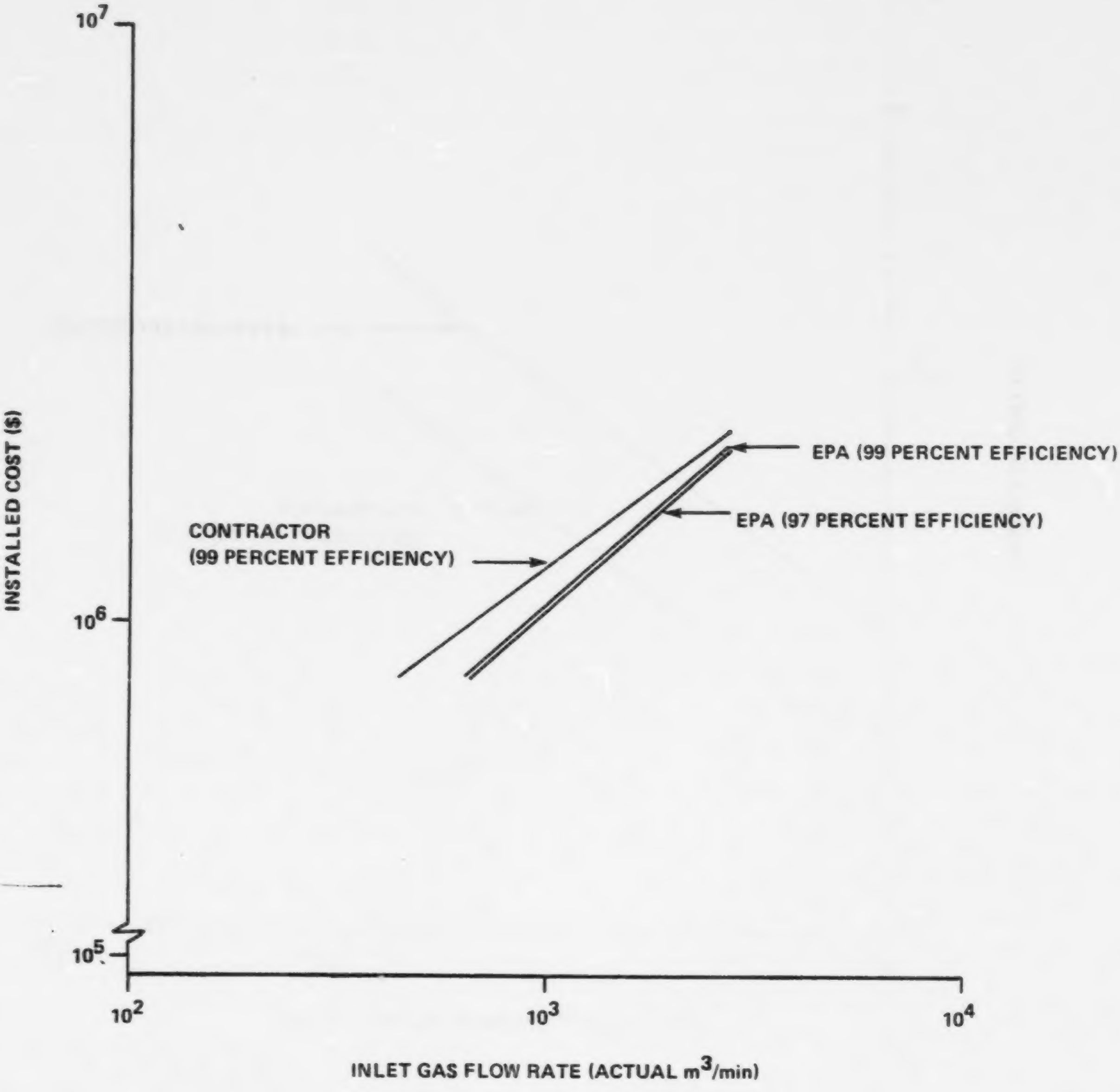

Figure 5-8. Installed costs of carbon adsorbers.

SOURCES: Lawson, J.F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride—Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

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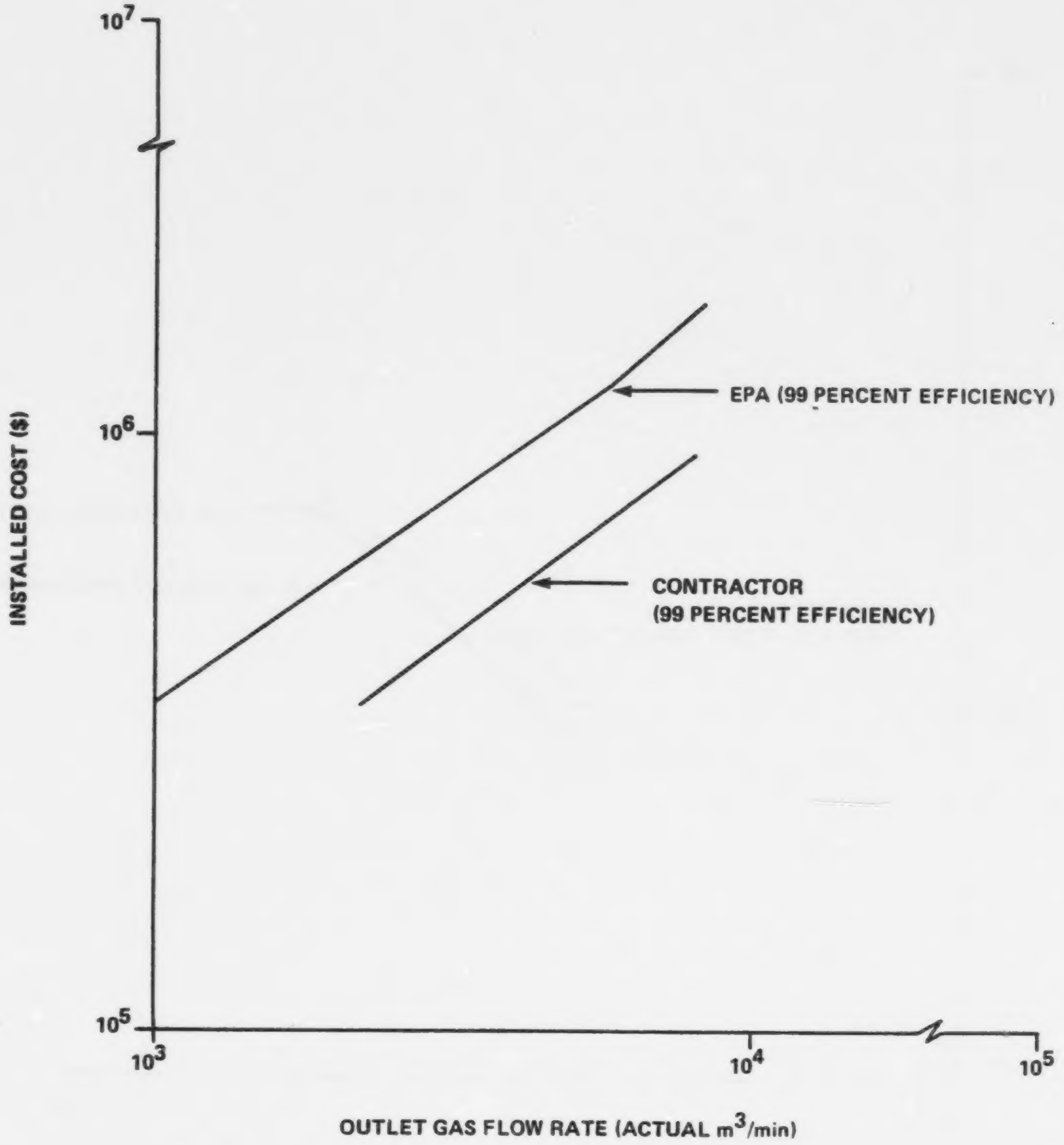

Figure 5-9. Installed costs of thermal incinerators with primary heat recovery.

SOURCE: Lawson, J.F. Emission Control Options for the Synthetic Organic Chemicals Manufacturing Industry Maleic Anhydride—Product Report. Hydroscience, Inc. (Prepared for Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency. Research Triangle Park, N.C.) EPA Contract Number 68-02-2577. March 1978.

Pruessner, R.D., and L.D. Broz. Air Pollution Control: Hydrocarbon Emission Reduction Systems. Chemical Engineering Progress. Petrotex Chemical Corp.:73. Houston, Texas. August 1977. p. 69-73.

5.4.2 Impact on Manufacturers

The direct economic effect of a benzene control standard would fall on MA manufacturers that use benzene as a feedstock. This impact, which depends upon the level of benzene control required, principally involves two areas--capital budget requirements and competition from domestic and foreign producers. The substitutes for MA are few and relatively unimportant in its primary markets, so little interindustry impact is expected.

5.4.2.1 Capital Budget Requirements. Two control levels for reducing benzene emissions are being considered: 97 percent and 99 percent reduction. At the 97-percent emission level, five plants will require no investment in controls because they already meet the standards. These plants are Koppers (Pa.), DENKA, Reichhold (N.J.)--which presently achieve this level of control--and Koppers (Ill.) and Amoco, which do not use benzene as a feedstock. Of these five plants, the Amoco and the two Koppers plants also meet the 99-percent emission level. Table 5-9 shows the capital budget requirements for plants requiring either carbon adsorption or incineration to achieve the standards being considered. The investment costs given in the table do not credit partial controls already in place on MA plants presently not meeting a possible standard. Thus, for plants presently meeting some level of benzene control, capital outlay assumptions for the 97- or 99-percent emission reduction standard may represent a "worst case."

In discussion of capital budget requirements, the main question is whether MA producers will have the financial resources from which to fund the initial capital investment. Funding can occur either externally, by sales of debt or equity, or internally, by using current capital budgets or allocating funds from the capital budgets of other divisions in the parent company. In addition to these financing techniques, costs may be passed through to consumers by raising the price of MA or other products produced by the firm to recover the capital expenses over a time period. Finally, although the method or combination of methods firms may employ to finance the additional capital expenditures cannot be predicted, it is likely that the expenditures will be treated as part of the parent company's overall capital budget.

In addition to the question of how the control costs might be financed is the question of whether such an investment is worthwhile. This decision

TABLE 5-9. ESTIMATED TOTAL INVESTMENT COST FOR ACHIEVING BENZENE EMISSION REDUCTION

Plant name & location	Type & level of control		Control equipment investment cost (10^3 dollars)	Monitoring equipment investment cost (10^3 dollars)	Total investment cost (10^3 dollars)
1. Ashland	CA	97%	1650	30	1680
Neal, W. Va.	CA	99%	1680	30	1710
	I	97%	1320	30	1350
	I	99%	1400	30	1430
2. Monsanto	CA	97%	2240	30	2270
St. Louis, Mo.	CA	99%	2270	30	2300
	I	97%	1730	30	1760
	I	99%	1860	30	1890
3. DENKA	CA	97%	0	UNK	0
Houston, Tex.	CA	99%	1440	30	1470
	I	97%	0	UNK	0
	I	99%	1220	30	1250
4. Reichhold	CA	97%	0	UNK	0
Elizabeth, N.J.	CA	99%	900	30	930
	I	97%	0	UNK	0
	I	99%	850	30	880
5. Reichhold	CA	97%	1250	30	1280
Morris, Ill.	CA	99%	1280	30	1310
	I	97%	1070	30	1100
	I	99%	1130	30	1160
6. Tenneco	CA	97%	780	30	810
Fords, N.J.	CA	99%	810	30	840
	I	97%	730	30	760
	I	99%	730	30	760
7. U.S. Steel	CA	97%	2250	30	2280
Neville Island, Pa.	CA	99%	2290	30	2350
	I	97%	1750	30	1780
	I	99%	1880	30	1910
8. Koppers	CA/I	97-99%	0	UNK	0
Bridgeville, Pa.					
9. Koppers	CA/I	97-99%	0	UNK	0
Chicago, Ill.					
10. Amoco	CA/I	97-99%	0	UNK	0
Joliet, Ill.					

NOTES: CA = carbon adsorption; I = incineration.

UNK indicates that the cost of monitoring equipment for those firms now meeting certain standard levels is unknown.

depends on the importance of MA to total company sales, capital availability, and the rate of return and risks relative to competing investment opportunities. Of the seven companies that could be affected by the alternative benzene standards being considered, five appear able to meet the additional capital budget requirements at both the 97- and 99-percent control levels. This assessment is based on analyses of physical and financial characteristics of individual plants as well as communications with industry. The other two companies may either choose not to or cannot finance such an investment.

One of these companies, DENKA, expressed concern that it could not finance additional benzene pollution controls.[15,16] The company has already made a major capital outlay for benzene controls within the past 4 years and is still recovering from this investment. It is a two-product company, with MA sales representing 33 percent of its total company sales. An investment of another 1.3 to 1.5 million dollars in new pollution control equipment could cause DENKA to go out of business, according to a company spokesperson.[15,16]

Although not confirmed by a company spokesperson, another possible candidate for closure is Tenneco. This opinion is based on several factors. First, Tenneco's plant, built in the early 1960's, is one of the oldest domestic MA plants. The number of remaining viable operating years at the plant, which may be few, will be a critical factor in an investment decision. Second, based on the economic analysis performed in this section, cost increments associated with a benzene standard would be highest for Tenneco under both control scenarios. Tenneco may therefore be at a cost disadvantage relative to its MA competitors. Third, it may be more profitable for the company to invest in products that are a more critical part of its product mix, or yield a higher return. Tenneco's MA sales[9] as a percentage of total company sales are the lowest of all affected companies, 0.05 percent (see Table 5-10). Finally, Tenneco's MA plant is located in a highly industralized area of New Jersey. Many chemical companies have already left this area and relocated in the South, where production and raw material costs are presently less expensive.

In summary, DENKA's possible closure response to a benzene standard is a stated company position and has been confirmed by an independent analysis.

TABLE 5-10. RATIO OF MA SALES TO PARENT COMPANY SALES[a]

	Sales of MA (1978 $ millions)	Parent company sales (1978 $ millions)	MA sales as percent of total parent company sales
1. Ashland	9.39	5,426	0.17%
2. Monsanto	13.16	5,019	0.26%
3. DENKA	NA[b]	NA	33.0%[c]
4. Reichhold	11.60	754	1.5%
5. Tenneco	4.07	8,762	0.05%
6. U.S. Steel	13.30	11,050	0.12%
7. Koppers	6.91	1,582	0.44%
8. Amoco	9.32	14,962	0.06%

[a] A 56-percent capacity figure was assumed in calculating MA sales for each company.

[b] NA = not available.

[c] Based on calculations by EPA.

It can be assigned a high degree of probability. Tenneco's closure possibility, however, has not been verified by a company spokesperson and therefore cannot be given as much confidence as the conclusions regarding DENKA's possible closure.

Table 5-11 compares the cost of the least expensive control option to total capital expenditures of the MA companies. Among the firms, Reichhold--which had the highest capital costs for control when expressed as a percentage of its capital budget--is expected to continue MA production. The capital budget requirement may affect Reichhold to a greater degree than the other companies (except for DENKA); nevertheless, Reichhold would probably fund needed control equipment and attempt to pass its increased costs to the consumer.

5.4.2.2 Intraindustry Competition. Intraindustry competition, prompted by a 97- or 99-percent benzene emissions standard, would chiefly involve the price differentials in MA merchant sales created by control costs being passed through to the consumer. Other potentially contributing factors are the interplay of transportation costs and the prospect of import competition. This section summarizes possible effects of the above factors on competition within the MA industry due to imposition of a benzene control standard.

5.4.2.3 Effect of Cost Pass-Through On Market Competition Due to Benzene Emissions Control. After funding initial capital to install controls, companies may be able to raise the price of MA to recover expenses incurred. The price increment each manufacturer would like to pass through to sales depends on the total investment costs and the quantity of product produced (i.e, the number of units through which price increments can be passed).

The MA industry is expanding production from approximately 56 percent toward the 100-percent production figure predicted for the post-1983 period.[2] Price increments due to cost pass-through should thus be viewed according to two production scenarios--56 percent and 100 percent capacity utilization. Tables 5-12 and 5-13 depict possible price increments of MA based on 100 percent cost pass-through and production capacities of 56 percent and 100 percent, respectively. In each case, the least expensive control option is employed, and costs are passed through according to both the production

TABLE 5-11. COMPARISON OF CONTROL COSTS TO TOTAL COMPANY CAPITAL EXPENDITURES

	1978 capital expenditures (10^6 dollars)	Control costs as percent of total company capital expenditures	
		97% control[a,b]	99% control[a,b]
Ashland	310	0.21	0.23
Monsanto	480	0.18	0.20
DENKA	NA	NA	NA
Reichhold	24	2.3	4.3
Tenneco	1,008	0.03	0.03
U.S. Steel	668	0.13	0.14
Koppers	144	0	0
Amoco	1,744	0	0

[a] Average annual capital cost assuming 2-year installation time and financing.

[b] Assumed the least expensive control between carbon adsorption and incineration.

SOURCE: Annual Reports and Form 10-K filed with the Securities and Exchange Commission Control Costs.

TABLE 5-12. POSSIBLE COST PASS-THROUGH UNDER CASE ASSUMPTION OF 56 PERCENT PRODUCTION CAPACITY[a]

	Company	97% benzene emissions control level			99% benzene emissiors control level		
		Increment (¢/kg)(¢/lb)	Total cost (¢/kg)(¢/lb)	% price increase	Increase (¢/kg)(¢/lb)	Total cost (¢/kg)(¢/lb)	% price increase
1.	Ashland	2.4 (1.1)	61.7 (28.0)	3.3	3.2 (1.5)	64.9 (29.4)	4.4
2.	Monsanto	2.3 (1.0)	64.0 (29.0)	3.1	3.1 (1.4)	64.8 (29.4)	4.2
3.	DENKA	0.0 (0.0)	61.7 (28.0)	0.0	3.3 (1.5)	65.0 (29.5)	4.5
4.	Reichhold (N.J.)	0.0 (0.0)	61.7 (28.0)	0.0	3.5 (1.6)	65.2 (29.6)	4.8
5.	Reichhold (Ill.)	2.6 (1.2)	64.3 (29.1)	3.5	3.3 (1.5)	65.0 (29.5)	4.5
6.	Tenneco	3.1 (1.4)	64.8 (29.4)	4.2	3.7 (1.7)	65.4 (29.6)	5.0
7.	U.S. Steel	2.3 (1.0)	64.0 (29.0)	3.1	3.1 (1.4)	64.8 (29.4)	4.2
8.	Koppers (Pa.)	0.0 (0.0)	61.7 (28.0)	0.0	0.0 (0.0)	61.7 (28.0)	0.0
9.	Koppers (Ill.)	0.0	61.7 (28.0)	0.0	0.0	61.7 (28.0)	0.0
10.	Amoco	0.0	61.7 (28.0)	0.0	0.0	0.0 (0.0)	0.0

[a] Costs are based on the expected low demand price of MA, 73.4¢/kg (33.3¢/lb).

TABLE 5-13. POSSIBLE COST PASS-THROUGH UNDER CASE ASSUMPTION OF 100 PERCENT PRODUCTION CAPACITY[a]

Company	97% benzene emissions control level			99% benzene emissions control level		
	Increment (¢/kg)(¢/lb)	Total cost (¢/kg)(¢/lb)	% cost increase	Increment (¢/kg)(¢/lb)	Total cost (¢/kg)(¢/lb)	% cost increase
1. Ashland	1.5 (0.7)	98.3 (44.6)	1.6	1.6 (0.7)	98.4 (44.6)	1.7
2. Monsanto	1.4 (0.6)	98.2 (44.5)	1.5	1.5 (0.7)	98.3 (44.6)	1.6
3. DENKA	0.0 (0.0)	96.8 (43.9)	0.0	1.6 (0.7)	98.4 (44.6)	1.7
4. Reichhold (N.J.)	0.0 (0.0)	96.8 (43.9)	0.0	1.8 (0.8)	98.6 (44.7)	1.9
5. Reichhold (Ill.)	1.6 (0.7)	98.4 (44.6)	1.7	1.7 (0.8)	98.5 (44.7)	1.8
6. Tenneco	1.8 (0.8)	98.6 (44.7)	1.9	2.0 (0.9)	98.8 (44.8)	2.1
7. U.S. Steel	1.4 (0.6)	98.2 (44.5)	1.5	1.5 (0.7)	98.3 (44.6)	1.6
8. Koppers (Pa.)	0.0 (0.0)	96.8 (43.9)	0.0	0.0 (0.0)	96.8 (43.9)	0.0
9. Koppers (Ill.)	0.0	96.8 (43.9)	0.0	0.0	96.8 (43.9)	0.0
10. Amoco	0.0	96.8 (43.9)	0.0	0.0	96.8 (43.9)	0.0

[a] Costs are based on the current actual price of MA, 96.8¢/kg (43.9¢/lb).

figures of each company and annualized capital control costs extended over a 10-year period. For the case of 56 percent capacity utilization, cost increments are based on the estimated actual price of molten MA (73.4¢/kg [33.3¢/lb]); for the case of 100 percent capacity utilization, cost increments are based on the current list price of MA (96.8¢/kg [43.9¢/lb]), a price reflecting expected MA prices in a high-demand market.*

It should be noted, in reference to Tables 5-12 and 5-13, that cost pass-through is being examined to show the effect of price differentials in the present MA market. Recognition of the effect of cost pass-through in the present market contributes to understanding the potential for its occurrence. This analysis is not meant to suggest that cost pass-through will happen; indeed, later analysis will show that passing costs through--unilaterally, in particular--is somewhat discouraged under present market conditions.

Passing costs through to the product can create price differentials between products of the different companies. Under each benzene control level--97 percent or 99 percent--some MA-producing companies are not required to install new benzene emissions controls. These companies incur no new expense and their products may remain at or near the same price. Significant price differentials, therefore, may result between the products of companies funding control costs and passing them through and those of companies not installing controls and/or choosing not to pass the expense through to the consumer.

Intraindustry competition caused by MA price differentials (resulting from cost pass-through of benzene controls) is sensitive to the operating rate of MA industry. Table 5-12 shows that, in a low-demand market, passing benzene control costs through fully can raise product prices by up to 4.5 percent in the 97-percent control case and by up to 4.8 percent in the 99-percent control case for an individual plant. In the high-demand, full-capacity utilization market (Table 5-13), potential cost increments would be lower, varying from 1.9 to 2.1 percent for the 97- and 99-percent control cases, respectively. Under both control and capacity utilization assumptions, Tenneco may be forced to raise prices the most, assuming it chooses to pass costs through.

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*Currently (August 1979), the list and actual price for MA are both 96.8¢/kg (43.9¢/lb). Due to the present tight market conditions, the actual price is the same as the list price. However, because market conditions fluctuate, it is likely MA companies will return to operating at lower rates if demand slackens. The actual price companies will receive under lower operating conditions cannot be determined. However, the difference between the actual and list price will probably be of similar magnitude to the differential that existed in 1978, a period of overcapacity in the MA market. Based on ratios, the actual price received in a low-demand market (56 percent operating capacity) has been calculated at 73.4¢/kg (33.3¢/lb).

It must be emphasized that, when intraindustry competition is viewed relative to possible MA price increases from cost pass-through, the major arena of competition involves merchant, rather than captive, markets. While costs may be passed through both markets, the greatest direct effect is seen in the highly competitive merchant market. Conversely, price increments in the captive MA market emerge only in the final product, having been diluted by costs of other constituents also necessary to produce the final product. The effect of this "dilution" is substantial (see Section 5.4.3). The following subsections, therefore, focus on the effects of cost differentials on MA merchant sales in both a low- and high-demand market.

5.4.2.3.1 Price differentials in a low-demand MA market. Raising prices in a low-demand market is not an auspicious prospect. Price elasticities governing the present MA market are simply not known. However, the MA industry has indicated that it would be reluctant to raise prices under low-demand market conditions; conversely, because most MA manufacturers are part of larger parent firms having other products, an alternative choice to total MA cost pass-through is to pass costs through partially or totally to other products. Some members of the MA industry have noted that they may employ this option, assuming the choice is made to install benzene emissions control if a standard were promulgated.[10,11,17,18,19]

Tables 5-14 and 5-15 further examine the effect of cost pass-through for two benzene control levels, based on the 56-percent production rate that approximates present MA operating conditions. Numbers in the table indicate the price differential (in cents per kilogram) between company and competitor, assuming 100 percent of cost pass-through. Negative numbers indicate a price advantage over a competitor, and the accompanying percent-

TABLE 5-14. DEPICTION OF POSSIBLE COMPETITIVE ADVANTAGES DUE TO COST PASS-THROUGH UNDER THE 56-PERCENT PRODUCTION CAPACITY ASSUMPTION

(97% Benzene Emissions Control Level)

Nameplate capacity (Gg)	Competitor	Company									
		Ashland	Monsanto	DENKA	Reichhold (N.J.)	Reichhold (Ill.)	Tenneco	U.S. Steel	Koppers (Pa.)	Koppers (Ill.)	Amoco
27	Ashland		-0.1 (-0.1)	-2.4 (-3.3)	-2.4 (-3.3)	0.2 (0.3)	0.7 (1.0)	-0.1 (-0.1)	-2.4 (-3.3)	-2.4 (-3.3)	-2.4 (-3.3)
48	Monsanto	0.1 (0.1)		-2.3 (-3.1)	-2.3 (-3.1)	0.3 (0.4)	0.8 (1.1)	0.0 (0.0)	-2.3 (-3.1)	-2.3 (-3.1)	-2.3 (-3.1)
23	DENKA	2.4 (3.3)	2.3 (3.1)		0.0 (0.0)	2.6 (3.5)	3.1 (4.2)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	0.0 (0.0)
14	Reichhold (N.J.)	2.4 (3.3)	2.3 (3.1)	0.0 (0.0)		2.6 (3.5)	3.1 (4.2)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	0.0 (0.0)
20	Reichhold (Ill.)	-0.2 (-0.3)	-0.3 (-0.4)	-2.6 (-3.5)	-2.6 (-3.5)		0.5 (0.7)	-0.3 (-0.4)	-2.6 (-3.5)	-2.6 (-3.5)	-2.6 (-3.5)
12	Tenneco	-0.7 (-1.0)	-0.8 (-1.1)	-3.1 (-4.2)	-3.1 (-4.2)	-0.5 (-0.7)		-0.8 (-1.1)	-3.1 (-4.2)	-3.1 (-4.2)	-3.1 (-4.2)
38	U.S. Steel	0.1 (0.1)	0.0 (0.0)	-2.3 (-3.1)	-2.3 (-3.1)	0.3 (0.4)	0.8 (1.1)		-2.3 (-3.1)	-2.3 (-3.1)	-2.3 (-3.1)
15	Koppers (Pa.)	2.4 (3.3)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	2.6 (3.5)	3.1 (4.2)	2.3 (3.1)		0.0 (0.0)	0.0 (0.0)
5	Koppers (Ill.)	2.4 (3.3)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	2.6 (3.5)	3.1 (4.2)	2.3 (3.1)	0.0 (0.0)		0.0 (0.0)
27	Amoco	2.4 (3.3)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	2.6 (3.5)	3.1 (4.2)	2.3 (3.1)	0.0 (0.0)	0.0 (0.0)	

NOTES: Top number in grid represents price differential between companies; negative sign indicates company has lower price than competitor. Number in parentheses depicts potential advantage in terms of cost differential expressed in percent (current list price of MA is basis: 73.4¢/kg [33.3¢/lb]).

TABLE 5-15. DEPICTION OF POSSIBLE COMPETITIVE ADVANTAGES DUE TO COST PASS-THROUGH UNDER THE 56-PERCENT PRODUCTION CAPACITY ASSUMPTION

(99% Benzene Emissions Control Level)

Name-plate capa-city (Gg)	Com-pet-itor	Company									
		Ashland	Monsanto	DENKA	Reichhold (N.J.)	Reichhold (Ill.)	Tenneco	U.S. Steel	Koppers (Pa.)	Koppers (Ill.)	Amoco
27	Ashland		-0.1 (-0.1)	0.1 (0.1)	0.3 (0.4)	0.1 (0.1)	0.5 (0.7)	-0.1 (-0.1)	-3.2 (-4.4)	-3.2 (-4.4)	-3.2 (-4.4)
48	Monsanto	0.1 (0.1)		0.2 (0.3)	0.4 (0.5)	0.2 (0.4)	0.6 (0.8)	0.0 (0.0)	-3.1 (-4.2)	-3.1 (-4.2)	-3.1 (-4.2)
23	DENKA	-0.1 (-0.1)	-0.2 (-0.3)		0.2 (0.3)	0.0 (0.0)	0.4 (0.5)	-0.2 (-0.3)	-3.3 (-4.5)	-3.3 (-4.5)	-3.3 (-4.5)
14	Reichhold (N.J.)	-0.3 (-0.4)	-0.4 (-0.5)	-0.2 (-0.3)		-0.2 (-0.3)	0.2 (0.3)	-0.4 (-0.5)	-3.5 (-4.8)	-3.5 (-4.8)	-3.5 (-4.8)
20	Reichhold (Ill.)	-0.1 (-0.1)	-0.2 (-0.3)	-0.0 (0.0)	0.2 (0.3)		0.4 (0.5)	0.2 (0.3)	-3.3 (-4.5)	-3.3 (-4.5)	-3.3 (-4.5)
12	Tenneco	-0.5 (-0.7)	-0.6 (-0.8)	-0.4 (-0.5)	-0.2 (-0.3)	-0.4 (-0.5)		-0.6 (-0.8)	-3.7 (-5.0)	-3.7 (-5.0)	-3.7 (-5.0)
38	U.S. Steel	0.1 (0.1)	0.0 (0.0)	0.2 (0.3)	0.4 (0.5)	0.2 (0.3)	0.6 (0.8)		-3.1 (-4.2)	-3.1 (-4.2)	-3.1 (-4.2)
15	Koppers (Pa.)	3.2 (4.4)	3.1 (4.2)	3.3 (4.5)	3.5 (4.8)	3.3 (4.5)	3.7 (5.0)	3.1 (4.2)		0.0 (0.0)	0.0 (0.0)
5	Koppers (Ill.)	3.2 (4.4)	3.1 (4.2)	3.3 (4.5)	3.5 (4.8)	3.3 (4.5)	3.7 (5.0)	3.1 (4.2)	0.0 (0.0)		0.0 (0.0)
27	Amoco	3.2 (4.4)	3.1 (4.2)	3.3 (4.5)	3.5 (4.8)	3.3 (4.5)	3.7 (5.0)	3.1 (4.2)	0.0 (0.0)	0.0 (0.0)	

NOTES: Top number in grid represents price differential between companies; negative sign indicates company has lower price than competitor. Number in parentheses depicts potential advantage in terms of cost differential expressed in percent (current list price of MA is basis: 73.4¢/kg [3.3¢/kg]).

age expresses that advantage in terms of a price differential divided by the assumed base price of MA prior to cost pass-through. Positive table numbers indicate that a company is marketing its MA at a higher price than its competitor. For example, Table 5-14 shows that Ashland's product costs 0.1¢ more per kilogram (0.05¢/lb) than Monsanto's; consequently, Monsanto may sell its product for less at a 0.1-percent differential over Ashland.

Under a 97-percent benzene control assumption, five companies would be required to install total control systems: Ashland, Monsanto, Reichhold (Ill.), U.S. Steel, and Tenneco. If these companies decided to pass costs through completely, they would be competing with other companies that had not installed new controls and whose MA prices could consequently be lower.

Table 5-14 shows that Tenneco, the producer with the largest potential price increment (3.1¢/kg [1.5¢/lb]), could be at a competitive disadvantage with other MA producers under a 97-percent benzene control standard. Of the competitors, all except Koppers (Ill.) have production capacities exceeding Tenneco's. Indeed, all have residual production capacity that could be used to capture part or all of Tenneco's market should price differentials encourage it. Similar situations, to a lesser extent, could exist for the other three companies required to install controls if they decide to pass costs through.

A 99-percent benzene emissions control would affect two more companies than the 97-percent standard (Table 5-15). Under the 99-percent control case, only the Amoco, Koppers (Pa.), and Koppers (Ill.) plants would not have to fund new control hardware. Again, Tenneco, assuming full cost pass through, would face the greatest price increase (3.7¢/kg [1.7¢/lb]), putting itself at a possible disadvantage with its competitors. Other companies would face similar situations, however, since the nonaffected companies-Koppers (Ill.), Koppers (Pa.), and Amoco--do not have enough residual capacity to capture all of the affected companies' markets. This situation would favor, to a degree, some cost pass-through, assuming the absence of increased import competition (see Section 5.4.3.1).

The effects of benzene emissions standards in a low-demand MA market can be summarized. In general, passing the costs of benzene controls through to MA may be difficult if MA demand is low. The market is quite competitive, and at least three companies--Koppers (Ill.), Koppers (Pa.), and Amoco--need not incur any new expense to meet a possible benzene emission standard. Because of these circumstances, MA companies financially affected by the standard may choose partial or total pass-through of benzene control costs to other company products as a means of offsetting the costs associated with required controls, while not placing themselves at a competitive disadvantage. Nevertheless, companies may not choose this option for a number of reasons.

Even in a depressed MA market situation, companies may choose instead to raise MA prices. Seven out of eight MA facilities will be required to install hardware under the 99-percent regulatory option (five for the 97-percent option), and each of these will want to pass control costs through to the consumer. If all or several companies with large capacities independently decide to pass costs through to MA, the market system may be able to sustain the higher MA prices. In this case, Amoco and Koppers would impose a limited threat of competition because neither company could produce MA in quantities sufficient to satisfy the entire domestic MA demand. Furthermore, Amoco, a large MA producer, has invested large sums of money into the research and development of its n-butane process. According to an Amoco spokesman, the company would readily raise prices to offset high R&D costs if a favorable market existed.[7] Thus, in the absence of import competition, higher MA prices may be a possible option for MA manufacturers even in a low-demand market.

5.4.2.3.2 Price differentials and the future MA market (at 100 percent capacity utilization). For at least the next 5 years, demand for the MA market is expected to continue to increase, corresponding to a 100-percent production rate of present MA capacity.[2] As this high-demand market is approached, increasing the price of MA will become more auspicious, and cost pass-through, thus, more possible. Furthermore, as the base price of MA increases to meet market demand (regardless of a benzene standard), price increments due to control cost pass-through become less significant. Table 5-13 shows that the highest MA price increments under both the 97- and 99-percent control level are 1.9 and 2.1 percent, respectively.

MA price differentials created from passing benzene control costs through to the consumer may not occur in an industry running at 100 percent production capacity. The largest producers are likely to be the price leaders as demand increases; thus, Monsanto and U.S. Steel may set prices in the high-demand market of 1983. Monsanto and U.S Steel, the largest MA producers, are both affected by the two possible benzene control level requirements. However, it is likely that as market demand increases to total MA production capacity, prices will rise to accommodate passing benzene control costs through with no concomitant loss in sales. (This assumes, of course, no import competition; see following sections.)

5.4.2.3.3 Effect of transportation on intraindustry impacts. Transportation charges can affect the delivered cost of MA. In 1978, costs for transporting MA by train averaged 0.005¢/kg/mi (0.002¢/lb/mi) and by truck, 0.01¢/kg/mi (0.005¢/lb/mi).[10] Thus, for 161 km (100 mi) of transport, the final cost of MA can increase from 0.5¢/kg to 1.0¢/kg. The current average transportation cost quoted is 3.86¢/kg, or between 300 and 400 mi of transport.[7] Thus, the final delivered price of MA may average 100.66¢/kg (45.63¢/lb) in today's market.

As MA prices increase, some companies may have an advantage over others in terms of transport credit. That is, considering MA prices of two competitors, a differential between the prices of their products could allow the lower priced product further transport distance before a "break-even" price with its more expensive competitor is reached. In an excess-capacity, low-demand market, such price differentials could allow one company to penetrate the regional market of another.

This possibility is illustrated in Tables 5-16 and 5-17 under different benzene emissions control levels. The tables assume regional markets exist for each company. Companies possibly breaking into a competitor's market area are shown horizontally, in the top row of the table. If a company from that row has a lower priced product than its competitor (shown in the vertical column), the company is given appropriate transport credit (in miles) equivalent to the price differential between the two companies' products. The distance between competing companies is also shown in the table, in the bottom number of each applicable grid. If a competitor has transport credit exceeding that distance, market penetration of a competitor is possible. However, it must be pointed out that many MA companies have an average market radius of between 300 and 400 miles.

TABLE 5-16. EFFECT OF TRANSPORTATION UNDER A 97-PERCENT CONTROL STANDARD

(Industry at 56% Production Capacity Utilization; Companies Possibly Entering Competitor's Market)

Name-plate capacity (Gg)	Competitor	Company									
		Ashland	Monsanto	DENKA	Reichhold (N.J.)	Reichhold (Ill.)	Tenneco	U.S. Steel	Koppers (Pa.)	Koppers (Ill.)	Amoco
27	Ashland		10/509	240/1278	240/524	NTC	NTC	10/213	240/163	240/457	240/442
48	Monsanto	NTC		230	230	NTC	NTC	NTC	230/563	230/289	230/274
23	DENKA	NTC	NTC		NTC	NTC	NTC	NTC	NTC	NTC	NTC
14	Reichhold (N.J.)	NTC	NTC	NTC		NTC	NTC	NTC	NTC	NTC	NTC
20	Reichhold (Ill.)	20/576	30/222	260/967	260/913		NTC	30/574	260/314	260/100	260/15
12	Tenneco	70/527	80/930	310/1558	310/50	50/915		80/341	310/366	310/752	310/767
38	U.S. Steel	NTC	NTC	230/1313	230/339	NTC	NTC		230/100	230/452	230/465
15	Koppers (Pa.)	NTC	NTC	NTC	NTC	NTC	NTC	NTC		NTC	NTC
5	Koppers (Ill.)	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC		NTC
27	Amoco	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC	

NOTES: NTC = no transportation credit.

A/B = numbers in table; where A is the number of miles designated, companies may be able to penetrate their competitor's market. These figures are based on price differentials between plants operating at 56 percent capacity with the 97-percent carbon absorption control.

A 0.01¢/kg/mi transportation cost was assumed in the calculations; B is the number of miles between designated companies.

TABLE 5-17. EFFECT OF TRANSPORTATION UNDER A 99-PERCENT CONTROL STANDARD

(Industry at 56% Production Capacity Utilization; Companies Possibly Entering Competitor's Market)

Nameplate capacity (Gg)	Competitor	Company									
		Ashland	Monsanto	DENKA	Reichhold (N.J.)	Reichhold (Ill.)	Tenneco	U.S. Steel	Koppers (Pa.)	Koppers (Ill.)	Amoco
27	Ashland		10/509	NTC	NTC	NTC	NTC	10/213	320/163	320/457	320/442
48	Monsanto	NTC		NTC	NTC	NTC	NTC	NTC	310/563	310/289	310/274
23	DENKA	10/1278	20/779		NTC	NTC	NTC	20/	330/1288	330/1067	330/1052
14	Reichhold (N.J.)	30/524	40/927	20/1555		20/	NTC	40/339	350/364	350/755	350/770
20	Reichhold (Ill.)	10/576	20/222	NTC	NTC		NTC	NTC	330/314	330/100	330/15
12	Tenneco	50/527	60/930	40/1558	20/50	40/915		60/341	370/366	370/752	370/767
38	U.S. Steel	NTC	NTC	NTC	NTC	NTC	NTC		310/100	310/452	310/465
15	Koppers (Pa.)	NTC	NTC	NTC	NTC	NTC	NTC	NTC		NTC	NTC
5	Koppers (Ill.)	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC		NTC
27	Amoco	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC	NTC	

NOTES: NTC = no transportation credit.

A/B = numbers in table; where A is the number of miles designated, companies may be able to penetrate their competitor's market. These figures are based on price differentials between plants operating at 56 percent capacity with the 99-percent thermal incineration control. A 0.01¢/kg/mi transportation cost was assumed in the calculations; B is the number of miles between designated companies.

Consequently, if between 600 and 800 miles are subtracted from the distance shown between two companies, it is evident that many companies already compete in common market regions. Another unknown factor within each plant's radius is distribution of its major market.

Tables 5-16 and 5-17 were calculated using the differential between companies at 56 percent capacity, assuming both a 97-percent and a 99-percent control level for benzene emissions, respectively. The transportation effects at 100 percent capacity utilization were not included because, at that utilization rate, companies would be selling to a high-demand market in which lack of buyers should not be an issue remedied by transport to other markets. At the 56-percent production capacity market, however, lack of buyers could be an issue.

Table 5-16 shows that, under a 97-percent benzene control level, the companies having the greatest potential for regional market penetration (due to MA price differentials) are DENKA, Reichhold (N.J.), U.S. Steel, Koppers, and Amoco. Of these, Amoco presents the greatest potential problem because of its large, residual production capacity, which can be used to capture part of a competitor's markets.

Table 5-17 shows that, under a 99-percent benzene emissions control, financially affected companies face potential regional market penetration by a greater number of competitors. In this situation, however, Amoco again presents the greatest threat because of its potentially large transport credit and its large residual production capacity.

Although both tables show that the competitive position of MA companies due to price differentials can be enhanced by transportation, this situation is unprecedented. Transportation costs currently limit the regional market of a producer to a radius of about 483 to 644 km (300 to 400 mi). Under a benzene emissions standard, assuming industries decide to pass control costs through, transportation may be used to sell MA beyond the traditional 483- to 644-km (300 to 400 mi) market radius and still meet competitive prices in certain cases, as illustrated in the tables. Whether companies will pass costs through and employ this method to achieve market penetration of a competitor cannot be determined. If MA producers elect to absorb pollution control cost and retain current prices, transportation costs could then serve to restrict market radii to below 483 or 644 km

(300 or 400 mi). In this case, regional market penetration of a competitor would not be profitable and would have low probability of occurring.

5.4.2.4 Effect of Imports. Imports have a small effect on the present MA market. In 1977, they accounted for only 2.3 percent of the United States MA demand.[12] The effect imports have on the future MA market is uncertain, however.

Maleic anhydride is produced in two forms: molten and briquette. Molten MA is the preferred form for large consumers, accounting for 90 percent of the domestic MA output.[11] The remaining 10 percent of MA is converted into briquettes, the form of all imported MA.[11] Because briquettes can be converted to the molten form only at a cost and because briquettes have a limited shelf life, imported MA competes primarily with domestic MA briquettes and not with the "molten market."

Imports could pose problems for United States MA briquette producers, although small quantities of imported MA have not interfered with domestic production. At times, imports have sold for 9¢ less per kilogram (4.1¢/lb) than domestic briquettes (which represent 10 percent of United States MA production), although imports now sell for only 2¢ less per kilogram (1.0¢/lb) than United States briquettes.[11] Prices could drop again, however. Foreign manufacturers are producing MA at below capacity and could readily expand if the demand existed. Moreover, imports may continue to compete with United States briquettes, particularly in the west coast market where arriving imports incur few additional overland freight charges.

Countering the potential of foreign competition are the facts that MA production costs are rising abroad and the possibility that foreign governments may stop subsidizing MA production. Furthermore, it should be reiterated that the degree of foreign competition is somewhat limited since the major arena of competition is briquette MA, which comprises only 10 percent of the domestic market.

Overall, the cost of benzene controls on domestic MA production may be somewhat exacerbated by foreign imports within the next 5 years. However, the United States has always met imported MA prices; if it continues to do so without incurring a loss until the post-1983 high-demand market is reached, imports should not present a problem.

5.4.2.5 Summary of Impact on Manufacturers. Tables 5-18 and 5-19 summarize the major impacts on the manufacturer arising from capital budget requirements and intraindustry competition under two benzene control levels--97 and 99 percent, respectively. Impacts of capital budget requirements indicate that Tenneco may be and DENKA will be most severely affected. Impacts of intraindustry competition indicate possible competition arising from benzene control cost pass-through in the future, high-demand market. Overall, the prospect of offsetting add-on benzene control costs and augmenting profits is enhanced in the future, high-demand market of 1983.

5.4.3 Effect on Product Prices

5.4.3.1 Cost Pass-Through to the Final Consumer. MA is purchased by manufacturers for use as an intermediate in the production of various end products. Should MA producers raise their product prices to compensate for pollution abatement costs, purchasers of MA will feel the greatest economic impact of these increments. It is therefore likely that MA users, like MA producers, will choose to raise prices of their products to offset higher production costs. Price increments of these products should not be as high as potential MA price increments because MA accounts for only a fraction of the wholesale price of each product. Thus, the impact of increased MA prices should be diluted in the final product.

MA is used as an intermediate in the production of several products, mainly polyester resins, agricultural chemicals, and fumaric acid. Tables 5-20, 5-21, and 5-22 show how the increased MA prices could affect the prices of these products. Product price increments depend on the percentage of MA in the product and the percentage of its wholesale price attributed to MA. These factors contribute to the varying increments associated with each product.

The predominant end use of MA is the production of unsaturated polyester resins (58 percent of MA demand) that go into reinforced plastic applications such as marine craft, building panels, automobiles, tanks, and pipes. Currently, these resins sell for 55¢ to $3.30/kg (25¢ to $1.50/lb). Should MA prices increase, polyester resins could rise from 0.2¢ to 0.5¢/kg (0.09¢ to 0.23¢/lb) for resins currently priced at 55¢/kg (25¢/lb) and from 1.8¢ to 3.0¢/kg (0.8¢ to 2.1¢/lb) for resins priced at $3.30/kg ($1.50/lb).*

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*Increments reflect the maximum range of increments that could occur using the worst price increases associated with the least expensive add-on control options at 97 and 99 percent benzene control.

TABLE 5-18. SUMMARY OF IMPACT OF 97 PERCENT BENZENE CONTROL LEVEL ON MA COMPANIES

		Intraindustry impact	
	Capital budget requirements	At 56% production capacity (present)	At 100% production capacity (1983)
Ashland	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition with Monsanto, U.S. Steel, Koppers, and Amoco	· Could pass costs through without sales loss
Monsanto	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from U.S. Steel, Koppers, and Amoco	· Could pass costs through without sales loss
DENKA	· Would not need to install a total control system · May not be able to meet capital budget requirements from internal resources · May choose to discontinue production	· Would have to pass costs through MA or other product line · Should have no new regional competitors	· Could pass costs through without sales loss
Reichhold, N.J.	· Would not need to install a total control system · Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers and Amoco	· Could pass costs through without sales loss
Reichhold, Ill.	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Ashland, U.S. Steel, Reichhold, Koppers, Monsanto, and Amoco	· Could pass costs through without sales loss
Tenneco	· Could fund costs but may discontinue MA production	· Would have highest production costs due to benzene control investment · Could face competition from Reichhold, Koppers, U.S. Steel, and Ashland	· Could pass costs through without sales loss
U.S. Steel	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers (Ill.) and Amoco	· Could pass costs through without sales loss · Possible price leader
Koppers, Pa.	· Would not need to install a total control system · Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Ashland or U.S. Steel	· Could pass costs through without sales loss
Koppers, Ill.	· Would not need to install controls	· Would not have to pass costs or raise prices	· Same as in present market · Would likely raise prices in response to demand
Amoco	· Would not need to install controls	· Would not have to pass costs or raise prices	· Same as in present market · Would likely raise prices in response to demand

TABLE 5-19. SUMMARY OF IMPACT OF 99 PERCENT BENZENE CONTROL LEVEL ON MA COMPANIES

		Intraindustry impact	
	Capital budget requirements	At 56% production capacity (present)	At 100% production capacity (1983)
Ashland	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers (Ill.) and Amoco	· Could pass costs through without sales loss
Monsanto	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers (Ill.) and Amoco	· Could pass costs through without sales loss · Possible price leader
DENKA	· May not be able to meet capital budget requirements from internal resources · May choose to discontinue production	· Would have to pass costs through MA or other product line · Should have no new regional competitors	· Could pass costs through without sales loss
Reichhold, N.J.	· Could fund costs and continued production	· Would have to pass costs through MA or other product line · Could face competition from Koppers (Ill.) and Amoco	· Could pass costs through without sales loss
Reichhold, Ill.	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers, (Ill.) and Amoco	· Could pass costs through without sales loss
Tenneco	· Could fund costs but may choose to discontinue production	· Would have highest production costs due to benzene control investment · Could face competition from U.S. Steel, Koppers (Ill.), and Amoco	· Could pass costs through without sales loss
U.S. Steel	· Could fund costs and continue production	· Would have to pass costs through MA or other product line · Could face competition from Koppers (Ill.) and Amoco	· Could pass costs through without sales loss · Possible price leader
Koppers, Pa.	· Would not need to install a total control system · Could fund costs and continue production	· Would have to pass costs through MA or other product line	· Could pass costs through without sales loss
Koppers, Ill.	· Would not need to install controls	· Would not have to pass costs or raise prices	· Same as in present market · Would likely raise prices in response to demand
Amoco	· Would not need to install controls	· Would not have to pass costs or raise prices	· Same as in present market · Would likely raise prices in response to demand

TABLE 5-20. PRICE INCREASES OF POLYESTER RESINS DUE TO INCREASED MA PRICES[a]

% control of benzene emissions	% production capacity of MA plant	Present wholesale price of polyester resins in ¢/kg (¢/lb)	% of wholesale price of polyester resins attributed to MA	Price increase of polyester resins in ¢/kg (¢/lb)	% increase of polyester resin prices	New wholesale price of polyester resins in ¢/kg (¢/lb)
97	56	55-330 (25-150)	15	0.4-2.5 (0.2-1.1)	0.7	55-333 (25-151)
	100			0.2-1.1 (0.1-0.5)	0.3	55-331 (25-150)
99	56	55-330 (25-150)	15	0.5-3.0 (0.2-1.4)	0.9	56-333 (25-151)
	100			0.2-1.4 (0.1-0.6)	0.4	55-331 (25-150)

[a] Price increments of polyester resins were calculated from the worst price increases associated with the "best-case" control options for MA.

TABLE 5-21. PRICE INCREASES OF FUMARIC ACID DUE TO INCREASED MA PRICES[a]

% control of benzene emissions	% production capacity of MA plant	Present wholesale price of fumaric acid in ¢/kg (¢/lb)	% of wholesale price of fumaric acid attributed to MA	Price increase of fumaric acid in ¢/kg (¢/lb)	% increase of fumaric acid prices	New wholesale price of fumaric acid in ¢/kg (¢/lb)
97	56	94 (43)	73	3.4 (1.5)	3.6	97 (44)
	100			1.6 (0.7)	1.7	96 (44)
99	56	94 (43)	73	4.1 (1.9)	4.4	98 (44)
	100			2.0 (0.9)	2.1	96 (44)

[a] Price increments of fumaric acid were calculated from the worst price increases associated with the "best-case" control options for MA.

TABLE 5-22. PRICE INCREASES OF MALATHION DUE TO INCREASED MA PRICES[a]

% control of benzene emissions	% production capacity of MA plant	Present wholesale price of malathion in ¢/kg (¢/lb)	% of wholesale price of malathion attributed to MA	Price increase of malathion in ¢/kg (¢/lb)	% increase of malathion prices	New wholesale price of malathion in ¢/kg (¢/lb)
97	56	242 (110)	11	1.3 (0.6)	0.5	243 (110)
	100			0.6 (0.3)	0.2	243 (110)
99	56	242 (110)	11	1.6 (0.7)	0.7	244 (111)
	100			0.8 (0.4)	0.3	243 (110)

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[a] Price increments of malathion were calculated from the worst price increases associated with the "best-case" control options for MA.

These increments are relatively small since MA accounts for only 15 percent of the total sales price of these resins; they represent only a 0.3- to 0.9-percent increase over the wholesale price of resins.

Agricultural chemicals, such as malathion, are the second largest users of MA (10 percent of MA demand). The current wholesale price of malathion is $2.42/kg ($1.10/lb),* and MA accounts for 11 percent of this price. Should MA prices go up, the price of malathion could rise from 0.6¢ to 1.6¢/kg (0.3¢ to 0.7¢/lb). Such an increment is relatively insignificant compared to the total cost of the product, representing an increase of only 0.2 to 0.7 percent over the present price of malathion.

Another market for MA is fumaric acid (5 percent of MA demand), which is primarily used as a food acidulant. MA accounts for approximately 73 percent of the total price of fumaric production. This large percentage is reflected in the maximum potential price increase of fumaric acid (1.6¢ to 4.1¢/kg [0.7¢ to 1.9¢/lb]). This increase represents a 1.70- to 4.40-percent increase over the current sale price of 94¢/kg (43¢/lb).

5.4.4 Employment and Balance of Trade

Imposing a benzene control standard could affect employment in the MA industry and could alter the present balance of trade. The MA industry presently employs approximately 330 individuals; under a 99-percent control standard, one manufacturer--DENKA--has indicated that capital expenditures for controls would prompt closure. This company represents 10 percent of the total domestic capacity. If it closes, approximately 33 workers could lose their jobs. Tenneco is also considered a closure candidate, although with less certainty than DENKA. If Tenneco did close, roughly 12 employees could lose their jobs.

Regarding balance of trade, foreign competition exacerbated by benzene control costs for U.S. producers could result in United States producers losing up to 12 percent of their market to imports. This number is based on the present 1977 import penetration of 2.3 percent in the total domestic market and the potential for United States briquette sales, the only form that is imported--10 percent of the total domestic market--to lose to foreign competition. The assumption of a 12-percent market loss to imports is highly conservative; more likely, imports will penetrate little or no more of the domestic market over the next 5 years (see Section 5.4.2.4).

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*Increments reflect the maximum range of increments that could occur using the worst price increases associated with the least expensive add-on control options at 97 and 99 percent benzene control.

5.4.5 Fifth-Year Impacts

This section summarizes the following aggregate economic impacts occurring 5 years after the standard is proposed:

- Total annualized costs,
- Net increase in national energy consumption, and
- Inflationary impact on the cost of MA. It is assumed that the plants will be running at full capacity at that time.

For the 97-percent regulatory option, total annualized costs would be about $2.1 million, including annualized capital costs and operation and maintenance costs of control and monitoring equipment. The additional consumption of energy with 50 percent heat recovery credit would be approximately 180 TJ/yr (29,000 bbl/yr). Furthermore, the price of MA would rise 1.2 percent over the current list price.* This is equivalent to about 1.0¢/kg (0.45¢/lb), assuming the full cost is passed through to the consumer.

For the 99-percent regulatory option, total annualized costs would be about $3.5 million. The additional energy consumption would be 540 TJ/yr (87,000 bbl/yr). MA prices would rise 1.5¢/kg (0.7¢/lb), or 1.7 percent of the current list price.

5.5 ECONOMIC IMPACTS OF USING n-BUTANE AS THE FEEDSTOCK AT NEW MALEIC ANHYDRIDE PLANTS

5.5.1 Introduction

This section assesses the economic impact of using n-butane as the alternate feedstock at new maleic anhydride facilities. The assessment includes impacts on domestic licensors of n-butane and benzene technology, on the availability and price of n-butane and benzene feedstock, and on the economic life of existing maleic anhydride facilities.

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*Derived by taking the increased total annualized cost (including monitoring costs) for the entire industry and dividing by total industry capacity (239 Gg). This approach was taken since it is not clear at this time which company, if any, will emerge as the price leader.

5.5.2 Impact on Licensors

There is only one domestic licensor of maleic anhydride technology, licensing both the benzene and n-butane processes.[11] Abroad, there are five licensing companies.[11] Of these five, only one licenses both the benzene- and n-butane-based technology. The other four license only the benzene-based process.

If n-butane were used as an alternate feedstock for new plants, it would be difficult to predict the effect on domestic producers. On one hand, it is likely the domestic licensor will maintain its foreign business because an EPA requirement would not affect usage of benzene-based technology abroad. However, the company's domestic business will depend on the competitive status of its n-butane and butene catalysts if benzene replacement were mandated.[20] Which companies are developing these catalysts--and the success of their experimental work--is presently unclear. Catalyst technology is usually a closely held company secret.

5.5.3 Impact on the Price and Availability of Feedstocks

At present, it is more economically favorable to use n-butane than benzene as a feedstock in a new facility. Use of the n-butane feedstock has been estimated to result in as much as a 7.3¢/kg (3.3¢/lb) MA cost reduction below benzene-based maleic anhydride.[9] And because the price of benzene has been rising at a faster rate than the price of n-butane, the cost differential between the two processes can be expected to widen.

Based on current trends in the benzene market, a tight benzene market has been predicted through 1990 (the time frame considered).[21] In the n-butane market, however, supplies should continue to be adequate to meet demand.[22]

A recent U.S. Department of Energy proposal to deregulate n-butane may, if promulgated, cause n-butane prices to rise. However, price increases that result from deregulation should not significantly change market conditions or the present makeup of consumers and end users.[23] Moreover, n-butane prices should experience smaller price increases than benzene, although a large increase in n-butane prices may result during the initial stages of deregulation.

The present benzene market is already tight. Benzene production in the United States is primarily refining-oriented, with availability and economics tied closely to gasoline and other fuels. One of the main sources of benzene production is from aromatics produced as a byproduct of the petroleum refining process. At present, little new refining capacity is being added in the industry because of an anticipated no-growth situation in domestic gasoline production. In addition, U.S. dependence on foreign oils has affected benzene production. Foreign oil is not as napthenic-rich as domestic crude and, therefore, has a lower aromatic concentration. As the aromatics concentration from which benzene is derived decreases, the economics of extraction become less attractive. Either less benzene is produced or higher benzene prices result.

At present, benzene consumed in U.S. MA production comprises 3.7 percent of the total benzene consumption.[21] This percentage is expected to increase to 4.4 percent by the year 1990.[21] However, this study does not account for EPA's possible regulation of new maleic anhydride plants.

n-Butane is primarily derived from natural gas as a coproduct of liquefied petroleum gas. Most n-butane produced in the United States is consumed directly at the refineries that produce it. The exact amount of refinery-consumed n-butane is difficult to calculate because refineries do not keep accurate accounts of the quantity of that n-butane that goes into the production of other refinery products.[22]

In spite of the difficulty associated with accurately calculating n-butane supply and demand at the refineries, figures on the n-butane unused by these refineries are available. Government statistics show at present an oversupply of n-butane on the market.[22] Because of this situation, the U.S. Department of Commerce recently considered the prospect of exporting domestic n-butane.[22]

The future availability of n-butane is linked directly to the future availability of natural gas and liquefied petroleum gas. At present, neither natural gas nor LPG production is expected to increase significantly over the next decade. However, even if increases are minimal, supplies of n-butane should adequately meet the domestic n-butane demand at new MA plants because refineries are expected to remain the prime users of n-butane and minimal increases are anticipated in their consumption pattern.[22]

5.5.4 Impact on the Economic Life of Existing Plants

A requirement of no detectable benzene emissions at new MA plants would have no impact on the economic life of existing plants. As mentioned in Section 5.2.3.2, use of n-butane feedstock has led to as much as a 7.3¢/kg (3.3¢/lb) MA cost reduction below benzene-based MA and could lead to a 25¢/kg (11¢/lb) differential. Because of the economic advantage associated with using n-butane feedstock, maleic anhydride manufacturers would opt to use the n-butane process in new plants regardless of EPA's requirement. Thus, any impacts that occur from new plants using n-butane would not be directly attributed to the EPA requirement.

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APPENDIX A

EVOLUTION OF THE BID

TABLE A-1. EVOLUTION OF THE BID

Date	Company, consultant, or agency	Nature of action	Location
June 8, 1977	EPA	Publication in the Federal Register of decision to list benzene as a hazardous air pollutant under Section 112 of the Clean Air Act.	
July 28-29, 1977	Reichhold Chemicals, Inc.	Plant visit.	Morris, Ill.
Sept. 28-29, 1977	Manufacturing Chemists Association Air Quality Committee	Meeting with engineers from EPA and Hydroscience, Inc., to review progress on the study of the maleic anhydride industry and control technology.	Washington, D.C.
October 20, 1977	Monsanto Chemical Intermediates Co.	Plant visit.	St. Louis, Mo.
October 25, 1977	Koppers Co., Inc.	Section 114 request for information.	Pittsburgh, Pa.
November 17, 1977	DENKA Chemical Corp.	Plant visit.	Houston, Tex.
December 21, 1977	Monsanto Chemical Intermediates Co.	Section 114 request for information.	St. Louis, Mo.
January 24, 1978	Amoco Chemicals Corp.	Plant visit.	Chicago, Ill.
February 2, 1978	Monsanto Chemical Intermediates Co.	Test data supplied to EPA.	St. Louis, Mo.
March 13-17, 1978	Reichhold Chemicals, Inc.	Carbon adsorption test by EPA.	Morris, Ill.
March 20-24, 1978	DENKA Chemical Corp.	Incinerator test by EPA.	Houston, Tex.
May 24, 1978	Reichhold Chemicals, Inc.	Plant visit and pretest survey.	Elizabeth, N.J.
June 2, 1978	EPA	Draft SSEIS mailed to working group.	Durham, N.C.
June 28, 1978	EPA	Meeting with working group to review comments on draft SSEIS.	Durham, N.C.
August 15, 1978	Koppers Co., Inc.	Section 114 request for information.	Pittsburgh, Pa.
August 23-24, 1978	EPA	NAPCTAC meeting.	Alexandria, Va.
October 31, 1978	Koppers Co., Inc.	Test data supplied to EPA.	Pittsburgh, Pa.
April 10, 1979	EPA	Steering committee meeting.	Washington, D.C.
April 25-26, 1979	DENKA Chemical Corp.	Plant visit.	Houston, Tex.
May 3, 1979	Ashland Chemical Corp.	Plant visit.	Neal, W. Va.
August 2, 1979	Tenneco	Section 114 request for information.	
September 1979	EPA	Steering committee meeting.	Washington, D.C.
October 1979	EPA	Delivery of AA Concurrence Package.	Washington, D.C.

APPENDIX B

INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

APPENDIX B

INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

Agency guidelines	Location within the BID
1. Background and description	
a. Process affected	The process to be affected is described in Section 1.2.1.
b. Industry affected	A description of the industry to be affected is given in Sections 1.1 and 5.2.
c. Availability of control technology	Information on the availability of control technology is given in Chapter 2.
2. Alternatives considered	
a. No action or postponing action	
· Environmental impacts	The environmental impacts of not implementing any standard are discussed in Section 4.1.
b. 97 percent control impact	
· Air pollution	The air pollution impacts of a 97-percent standard are discussed in Sections 4.1.2 and 4.1.4.
· Water pollution	The water pollution impact of a 97-percent standard is discussed in Section 4.2.
· Solid waste disposal	The impact of a 97-percent control standard on solid waste disposal is discussed in Section 4.3.
· Energy impact	The energy impacts of a 97-percent control standard are discussed in Sections 4.4.2 and 4.4.4.
· Economic impact	The economic impacts of a 97-percent control standard are discussed in Sections 5.3 and 5.4.

Agency guidelines	Location within the BID
c. 99 percent control impact	
· Air pollution	The air pollution impacts of a 99-percent control standard are discussed in Sections 4.1.1 and 4.1.3.
· Water pollution	The water pollution impacts of a 99-percent control standard are discussed in Section 4.2.
· Solid waste disposal	The solid waste disposal impacts of a 99-percent control standard are discussed in Section 4.3.
· Energy	The energy impacts of a 99-percent control standard are discussed in Sections 4.4.1 and 4.4.3.
· Economic	The economic impacts of a 99-percent control standard are discussed in Sections 5.3 and 5.4.
d. Conversion to n-butane for existing plants	
· Feasibility	The feasibility of converting existing plants to the n-butane process is discussed in Sections 1.2.2 and 2.1.4.
· Environmental impact	The environmental impacts of the n-butane process are discussed in Section 4.1.5.
e. n-Butane use for new plants	
· Air pollution impact	The air pollution impacts of requiring new plants to use n-butane are discussed in Section 4.1.5.
· Water pollution impact	The water pollution impacts of requiring new plants to use n-butane are discussed in Section 4.2.
· Solid waste disposal	The solid waste disposal impacts of requiring new plants to use n-butane are discussed in Section 4.3.

Agency guidelines	Location within the BID
· Economic impact	The economic impacts of requiring new plants to use n-butane are discussed in Section 5.5.
· Energy impact	The energy impacts of requiring new plants to use n-butane are discussed in Section 4.4.5.

TABLE B-1. MATRIX OF ENVIRONMENTAL IMPACTS OF ALTERNATIVE CONTROL SYSTEMS

Alternative	Impact on benzene	Impact on HC	Other air impacts	Water impact	Solid waste impact	Energy impact	Space impact	Noise impact	Economic impact
A. 97% control	+3	+3	-1	-1	-1	-1	-1	-1	-2
B. 99% control	+4	+4	-1	-1	-1	-3	-1	-1	-3
A. Substitution of feedstock, such as n-butane (new plants only)	+5	-2	-1	-1	0	0	0	0	+2
Delayed standard	-2	-2							0
No standard	-3	-3							0

KEY	+	Beneficial impact	0	No impact
	-	Adverse impact	1	Negligible impact
			2	Small impact
			3	Moderate impact
			4	Large impact
			5	Larger impact

APPENDIX C

EMISSION SOURCE TEST DATA

APPENDIX C

EMISSION SOURCE TEST DATA

C.1 INTRODUCTION

The purpose of this appendix is to present and summarize data gathered during the development of a standard for benzene emissions from the production of maleic anhydride. The facilities tested are described and the source testing methods are identified. Any reference to commercial products and processes by name does not constitute endorsement by the U.S. Environmental Protection Agency.

C.2 SUMMARY

Two separate yet similar facilities producing maleic anhydride were tested to determine percent reduction of emissions. Each plant was tested by a different contractor. Pollutants analyzed and procedures used were:

Benzene	Draft EPA Method
Total hydrocarbons (THC)	Draft EPA Method
*Carbon dioxide	EPA Method 10/Orsat
*Oxygen	EPA Method 10/Orsat
*Carbon monoxide	EPA Method 10/Orsat
*Methane	Draft EPA Method
*Ethane	Draft EPA Method
Total organic acids (TOA)	†LAAPCD Method
Total aldehydes	LAAPCD Method
Formaldehyde	LAAPCD Method
Temperature	Thermocouple
Duct pressure	EPA Method 2
Volumetric flow rate	EPA Method 2
NO_x	EPA Method 7

Although production facilities are similar in the two plants, the means of emissions control are different. Plant A controls emissions by use

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*Data not presented in text.

†LAAPCD--Los Angeles Air Pollution Control District.

of a carbon adsorption system, while Plant B uses an incinerator. The available data indicate that incineration provides slightly higher removal efficiencies for benzene and total hydrocarbons, while carbon adsorption is acceptable and is more consistent in removing emissions.

C.3 DESCRIPTION OF FACILITIES

C.3.1 Plant A

This plant uses a single-train, multiple reactor process with a capacity of 20,000 Mg/yr of maleic anhydride. At the time samples for this study were taken, the plant was operating at about 40 percent of its design capacity.

Maleic anhydride is produced at Plant A (Table C-1) by the vapor phase oxidation of benzene in a tubular reactor. The resulting reactor exhaust passes through a series of switch condensers that are first cooled to freeze the maleic anhydride from solution and later heated to melt the maleic anhydride for pumping to storage. After the freezing process, the remaining exhaust gas enters the product recovery absorber, which scrubs the exhaust with water or aqueous maleic acid. The liquid effluent from the absorber is about 40 percent, by weight, maleic acid. Vented emissions from the absorber are directed to the carbon adsorption system. Essentially all process emissions exit through the product recovery absorber.

The single inlet sampling port is located in the 1.07-m I.D. duct leading into the three carbon adsorption units. The outlet sampling port is located in the 1.07-m I.D. stainless steel exhaust stack, 11.6 m from the bottom of its 29.3-m height.

Integrated gas samples were obtained in Tedlar® bags (0.113 m^3 [4 ft^3]), heated to a temperature of 49° to 74° C (120° to 165° F). Sampling commenced at the onset of a 1-hour desorption cycle and lasted for three full cycles (3 hours).

Benzene and total hydrocarbon concentrations were determined in the field by gas chromotography with flame ionization detection. Methane and ethane concentrations were not determined because of inadequate resolution by retention time. Total hydrocarbons were determined as propane.

Total aldehydes were determined using the LAAPCD Method, involving sample reaction with sodium bisulfate, pH adjustment freeing bisulfate ions in an amount equivalent to the aldehydes present, and titration with a

TABLE C-1: PLANT A--EMISSIONS SUMMARY

Sample designation		Inlet	Outlet	% reduction
	Benzene:			
A	Concentration (ppm)	861	81.4	90.5
	Emission rate (kg/hr)	83.2	7.87	
B	Concentration (ppm)	977	63.9	93.5
	Emission rate (kg/hr)	92.3	6.04	
C	Concentration (ppm)	915	42.2	95.4
	Emission rate (kg/hr)	86.9	4.01	
	Total hydrocarbon concentration (as propane):			
A	Concentration (ppm)	1,610	147	90.9
	Emission rate (kg/hr)	87.9	8.02	
B	Concentration (ppm)	1,600	110	93.1
	Emission rate (kg/hr)	85.3	5.87	
C	Concentration (ppm)	1,360	67.3	95.1
	Emission rate (kg/hr)	73.0	3.61	
	Total organic acids (as maleic acid):			
A	Concentration (ppm)	17.6	0.39	97.8
	Emission rate (kg/hr)	2.53	0.06	
B	Concentration (ppm)	<4.00	<0.24	94.0
	Emission rate (kg/hr)	<0.56	<0.03	
C	Concentration (ppm)	6.58	1.16	82.4
	Emission rate (kg/hr)	0.93	0.16	
	Total aldehydes (as formaldehyde):			
A	Concentration (ppm)	88.0	<9.21[a], 3.6[b]	N.A.[a,c]
	Emission rate (kg/hr)	3.3	<0.34[a], 0.13[b]	96.0[b]
B	Concentration (ppm)	112	13.0[a], 2.4[b]	88.4[a]
	Emission rate (kg/hr)	4.1	0.47[a], 0.09[b]	97.9[b]
C	Concentration (ppm)	74.8	109[a], 2.5[b]	N.A.[a,d], 96.7[b]
	Emission rate (kg/hr)	2.7	4.0[3a], 0.09[b]	
	Formaldehyde:			
A	Concentration (ppm)	70.9	1.8[a], 1.8[b]	97.5[a]
	Emission rate (kg/hr)	2.6	0.07[a], 0.07[b]	97.5[b]
B	Concentration (ppm)	85.5	1.8[a], 1.4[b]	97.9[a]
	Emission rate (kg/hr)	3.1	0.06[a], 0.05[b]	98.4[b]
C	Concentration (ppm)	54.2	1.1[a], 1.3[b]	97.9[a]
	Emission rate (kg/hr)	2.0	0.04[a], 0.05[b]	97.6[b]

[a] Flask analysis of bag sample.

[b] Sample from continuous isokinetic impinger sampling train.

[c] Not available due to "less than" values of outlet.

[d] Not available due to outlet concentration being greater than inlet concentration.

standard iodine solution. Formaldehyde is determined by sample reaction with sodium bisulfate, reaction with chromotropic acid, and colorimetric determination of a uniquely colored compound formed in the second reaction. Samples for formaldehyde and total aldehydes were taken from the bag samples and from a continuous isokinetic impinger sampling train. Both results are presented.

Analysis for Total Organic Acids (TOA) was conducted in accordance with the LAAPCD Air Pollution Source Testing Manual, November 1963. This method involves "field fixing" samples with sodium hydroxide and transfer to the lab for ether extraction of organic acids and titration with a standard base.

The EPA Method 3 (Orsat) was used for determining CO_2, O_2, and CO.

For Plant A, in addition to analysis of the gas stream at the inlet and outlet of the emission control system, water samples from the drain to the product recovery absorber were analyzed for benzene, TOA, formaldehyde, and total aldehydes.

C.3.2 Plant B

This plant has a maximum production capacity of 23,000 Mg/yr of maleic anhydride. At the time samples for this study were taken, the plant was operating at about 70 percent of its design capacity. Plant personnel did not think the lower production rate would seriously affect the validity of the results.

Maleic anhydride is produced at Plant B (Table C-2) by the vapor phase oxidation of benzene in a tubular reactor. The resulting reactor exhaust gas passes through a partial condenser that separates out a portion of the crude maleic anhydride. The remaining exhaust gas enters the product recovery absorber, which scrubs the exhaust with water or aqueous maleic acid and produces an aqueous solution containing about 40 percent, by weight, maleic acid. The exhaust gas from the absorber is directed to an incinerator. Essentially all process emissions exit through the product recovery absorber.

A single inlet sampling port is located in the 0.91-m ID inlet duct to the incinerator. An outlet sampling port is located at each of eight 7.5-cm ID incinerator outlet ducts. The outlet ports were located downstream of any physical disturbances.

TABLE C-2. PLANT B--EMISSIONS SUMMARY

Sample designation		Inlet	Outlet	% reduction
	Benzene:			
A	Concentration (ppm)	780	11.1	98.6
	Emission rate (kg/hr)	154	2.2	
B	Concentration (ppm)	820	11.8	98.6
	Emission rate (kg/hr)	161	2.4	
C	Concentration (ppm)	940	14.4	98.5
	Emission rate (kg/hr)	185	2.9	
	Total hydrocarbon concentration (as propane):			
A	Concentration (ppm)	1,520	24.3	98.4
	Emission rate (kg/hr)	169	2.7	
B	Concentration (ppm)	1,880	23.6	98.7
	Emission rate (kg/hr)	209	2.8	
C	Concentration (ppm)	2,090	26.5	98.7
	Emission rate (kg/hr)	232	3.0	
	Total organic acids (as maleic acid):			
A	Concentration (mg/m^3)	153	362	N.A.[a]
	Emission rate (kg/hr)	9.3	21.8	
B	Concentration (mg/m^3)	208	89	54.0
	Emission rate (kg/hr)	12.6	5.8	
C	Concentration (mg/m^3)	281.1	57	78.8
	Emission rate (kg/hr)	17.0	3.6	
	Total aldehydes (as formaldehyde):			
A	Concentration (mg/m^3)	33.9	4.7	85.0
	Emission rate (kg/hr)	2.0	0.3	
B	Concentration (mg/m^3)	49.7	6.2	86.7
	Emission rate (kg/hr)	3.0	0.4	
C	Concentration (mg/m^3)	90.1	4.7	94.4
	Emission rate (kg/hr)	5.4	0.3	
	Formaldehyde:			
A	Concentration (mg/m^3)	11.0	0.60	94.0
	Emission rate (kg/hr)	0.67	0.04	
B	Concentration (mg/m^3)	21.8	0.00	100
	Emission rate (kg/hr)	1.32	0.00	
C	Concentration (mg/m^3)	47.8	0.60	98.6
	Emission rate (kg/hr)	2.90	0.04	
	Oxides of nitrogen:			
A	Concentration (mg/m^3)	N.D.	15.7	N.D.
	Emission rate (kg/hr)	N.D.	1.0	
B	Concentration (mg/m^3)	N.D.	17.4	N.D.
	Emission rate (kg/hr)	N.D.	1.1	
C	Concentration (mg/m^3)	N.D.	13.6	N.D.
	Emission rate (kg/hr)	N.D.	0.9	

[a] Not available due to outlet concentration being higher than inlet concentration.

NOTE: N.D. = not determined.

Gas samples were obtained according to the EPA draft method for benzene, in 70-L (2.5 cf) Mylar® bags, heated to 66° C (151° F). Total hydrocarbons, benzene, methane, and ethane were determined by gas chromatography, with flame ionization detection. Total hydrocarbons were determined as propane. Carbon dioxide, oxygen, and carbon monoxide were analyzed by nondispersive infrared spectroscopy, using Orsat analysis to determine percentages.

Total organic acids in the samples were analyzed according to the LAAPCD Method, which involves bubbling the sampled gas through a dilute sodium hydroxide solution, acidification, ether extraction of the organic acids, and titration with a standard base. The LAAPCD Method employed for total aldehydes entails sample reaction with sodium bisulfate, pH adjustment freeing bisulfate ions, and titration with a standard iodine solution. Formaldehyde is determined by sample reaction with sodium bisulfate, reaction with chromotropic acid, and colorimetric determination of a uniquely colored compound formed in the second reaction.

Oxides of nitrogen (NO_x) were determined according to EPA Reference Method 7, employing colorimetric analysis of a complex ion formed by reaction of organic acids with the NO_x. Duct temperature was obtained by thermocouple, and duct pressure by water manometer, according to EPA Reference Method 2.

APPENDIX D

EMISSION MEASUREMENT AND CONTINUOUS MONITORING

APPENDIX D

EMISSION MEASUREMENT AND CONTINUOUS MONITORING

D.1 EMISSION MEASUREMENT METHODS

D.1.1 General Background

For stack sampling purposes, benzene will, except in the case of systems handling pure benzene, exist in the presence of other organics. Accordingly, methods for benzene analysis consist first of separating the benzene from other organics, followed by measuring the quantity of benzene with a flame ionization detector. However, among various stack-testing groups concerned with measuring benzene, there is no uniformity of procedures. Problems of uniformity could exist in the following areas:

- Sample collection,
- Sample introduction to gas chromatograph,
- Chromatographic column and associated operating parameters, and
- Chromatograph calibration.

Two possible approaches for benzene sample collection are grab samples and integrated samples. Because benzene emission concentrations may vary considerably during a short period, the integrated sample approach is more advantageous since emission fluctuations due to process variations are automatically averaged. In addition, the integrated approach reduces the number of samples that need to be analyzed. For integrated samples, both tubes containing activated charcoal and Tedlar® bags have been used. However, charcoal sampling tubes were basically designed for sampling ambient concentration levels of organics. Because source effluent concentrations are expected to be higher (particularly since organics other than benzene could be present), uncertainty would be involved in predicting sample breakthrough or sample termination. Bag samples would also offer the best potential for precision because no intermediate sample recovery step would be involved.

Based on the above considerations, collecting an integrated sample in Tedlar® bags appears to be the best alternative, a conclusion shared by an EPA-funded study whose purpose was to propose a general measurement technique for gaseous organic emissions.[1] Another study of benzene stability in Tedlar® bags was undertaken to confirm the soundness of this approach.[2] Because this study showed no significant deterioration of benzene over a period of 4 days, the integrated bag technique was deemed suitable. However, anyone preferring to use activated charcoal tubes has this option, provided that efficiency equal to or better than the bag technique can be demonstrated and procedures to protect the integrity of the sampling technique are followed.

A collected gas sample can be introduced to a gas chromatograph either by using a gas-tight syringe or an automated sample loop. The latter approach was selected for the reference method because it has less potential for leakage and provides a more reproducible sample volume.

Several columns are mentioned in the literature that may be suitable for separating benzene from other gases.[3,4] Most notable are 1, 2, 3-tris (2-cyanoethoxy) propane for separating aromatics from aliphatics and Bentone 34 for separating aromatics. A program was undertaken to establish whether various organics associated with benzene in stack emissions interfered with the benzene peaks from the two columns.[5,6] The study revealed the former column to be suitable for analyzing benzene in gasoline vapors and the latter column to be suitable for analyzing benzene emissions from maleic anhydride plants. It should be noted that selection of these two columns for inclusion in Method 110 does not mean that some other column(s) may not work equally well. In fact, the method has a provision for using other columns.

Calibration has been accomplished by two techniques, the most common being the use of cylinder standards. The second technique involves injecting known quantities of 99 mole percent pure benzene into Tedlar® bags as they are being filled with known volumes of nitrogen. The second technique has been found to produce equally acceptable results; both are included in Method 110.

D.1.2 Field Testing Experience

Based on the study of benzene stability in Tedlar® bags, of possible interferences by various process-associated gases, and of calibration methods, and as a result of a field study and tests conducted at sources of benzene emissions, a new draft of Method 110 for determining compliance with benzene standards or NESHAPS was prepared. This method is the same as the originally investigated method, except that the audit procedure has been refined and an appendix has been added to help verify benzene peak resolution.

Two maleic anhydride manufacturing plants were tested during the development test program. One of these plants employed a carbon adsorption system, and the other employed a thermal oxidizer (incinerator) to control organic emissions. Method 110 was used to collect and analyze for benzene emissions. The SP 1200/Bentone 34 gas chromatographic column described in the method was used to resolve the benzene. No major deviations from Method 110 were required. At the plant employing the carbon adsorption system, a liquid dropout was required to prevent intermittent entrained liquid from being introduced into the integrated bag sample. This liquid entrainment, caused when an undried steam-desorbed carbon bed was reintroduced into the control system, occurred 5 percent of the sampling time. A rotameter was also installed at the inlet to the integrated bag sample at this plant because the rigid container housing the bag sample could not be adequately sealed. Neither of these two deviations are considered to have affected data validity.

The sampling lines and bags were maintained at or slightly above the source temperature during collection and analysis to prevent condensation of organics that would normally be a vapor at the source temperature.

Organic acid and aldehyde emission data were also collected during the test program. Data were collected to adequately determine emissions in terms of concentration, mass rate, and control system mass removal efficiency.

D.2 CONTINUOUS MONITORING

No emission-monitoring instrumentation, data acquisition, or data processing equipment have been identified for measuring benzene from maleic anhydride plant stack gases. However, the U.S. Environmental Protection

Agency (EPA) has only recently begun to explore the development of specifications for a system for benzene monitoring; such a system, which would employ a package of individual, commercially available items, is considered feasible.

For a chromatographic system that measures benzene concentration, the installed cost of the chromatograph and its auxiliaries, which include gas chromatograph with dual-flame detector, automatic gas sampling valve, air sampler, post-run calculator, and gas regulators, is $35,000. This figure would increase by approximately $10,000 for the additional hardware necessary to report a benzene mass emissions rate in terms of benzene feedstock. Depending on the operating factor, the direct operating cost varies from about $1,200 to $1,400 per year.

D.3 EMISSION TEST METHODS

The recommended emission test method for determining benzene emission concentrations at maleic anhydride plants is Method 110. The method uses the Method 106 train for sampling and a gas chromatograph/flame ionization detector equipped with a column selected for separation of benzene from the other organics present for analysis.

Subpart A of 40 CFR 61 requires that facilities subject to National Emission Standards for Hazardous Air Pollutants be constructed to provide sampling ports adequate for the applicable test methods, and platforms, access, and utilities necessary to perform testing at those ports.

Assuming that the test location is near the analytical laboratory and that sample collection and analytical equipment are available, the cost of field collection, laboratory analysis, and reporting of benzene emissions in triplicate from a single stack is estimated to be $2,500 to $3,500 for an emission test effort. This figure assumes a cost of $25 per person-hour. This amount would be reduced by approximately 50 percent per stack if several stacks were tested.

If the plant established in-house sampling capabilities and conducted its own tests and/or analyses, the cost per person-hour could be less.

D.4 REFERENCES

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4. Current Peaks. 10:1. Carle Instruments, Inc. Fullerton, Calif. 1977.

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APPENDIX E

METHODOLOGY FOR ESTIMATING MORTALITY AND LIFETIME RISK FROM EXPOSURE TO BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS

APPENDIX E

METHODOLOGY FOR ESTIMATING MORTALITY AND LIFETIME RISK FROM EXPOSURE TO BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS

E.1 INTRODUCTION

The purpose of this appendix is to describe the methodology used in estimating leukemia mortality and lifetime risk attributable to population exposure to benzene emissions from maleic anhydride manufacturing plants. The appendix is presented in three parts:

- Part E.2, Summary and Overview of Health Effects, summarizes and references reported health effects from benzene exposure. The major reported health effect is leukemia. Mortalities cited in the BID include only the estimated leukemia cases attributable to exposure to benzene emissions from existing maleic anhydride plants although other, sometimes fatal, effects are known to result from benzene exposure.

- Part E.3, Population Density Around Maleic Anhydride Plants, describes the method used to estimate the population at risk; i.e., persons residing within 20 km of existing maleic anhydride plants.

- Part E.4, Population Exposures, Risks, and Mortalities, describes the methodology for estimating benzene emissions from a model plant, calculating expected population exposures, and estimating leukemia deaths attributable to benzene emissions from the eight existing U.S. plants.

E.2 SUMMARY AND OVERVIEW OF HEALTH EFFECTS

E.2.1 Health Effects Associated with Benzene Exposure

A large number of occupational studies over the past 50 years have provided evidence of severe health effects in humans from prolonged inhalation exposure to benzene. Some 300 studies[1] of the health effects of benzene have recently been reviewed and analyzed in terms of application to low-level ambient benzene exposures that might occur in a population residing near a source of benzene emissions.

The reviewers concluded that benzene exposure by inhalation is strongly implicated in three pathological conditions that may be of public health concern at environmental exposure levels:

- Leukemia (a cancer of the blood-forming system),
- Cytopenia (decreased levels of one or more of the formed elements in the circulating blood), and
- Chromosomal aberrations.

Leukemia is a neoplastic proliferation and accumulation of white blood cells in blood and bone marrow. The four main types are acute and chronic myelogenous leukemia and acute and chronic lymphocytic leukemia. The causal relationship between benzene exposure and acute myelogenous leukemia and its variants in humans appears established beyond reasonable doubt.[1]

The term "pancytopenia" refers to diminution of all formed elements of the blood and includes the individual cytopenias: anemia, leukopenia, thrombocytopenia, and aplastic anemia. In mild cases, symptoms of pancytopenia are such nonspecific complaints as lassitude, dizziness, malaise, and shortness of breath. In severe cases, hemorrhage may be observed, and death may occasionally occur because of hemorrhage or massive infection. Patients with pancytopenia may subsequently develop fatal, acute leukemia.

Chromosomal aberrations include chromosome breakage and rearrangement and the presence of abnormal cells. These aberrations may continue for long periods in hematopoietic and lymphoid cells. The health significance of these aberrations is not fully understood. However, aberrant cells have been observed in individuals exposed to benzene who have later developed leukemia. Some types of chromosomal aberrations may be inheritable.

In one study too recent to include in the review previously cited,[1] workers exposed to 2.1 ppm benzene for 4 years showed a statistically significant increase in chromosomal aberrations (as high as tenfold) over those in unexposed controls.[2]

The review[1] concluded that man may be the only species yet observed to be susceptible to benzene-induced leukemia. Evidence for production of leukemia in animals by benzene injection was considered nonconclusive.

Moreover, benzene exposure by oral dosing, skin painting, or inhalation has not been shown to produce leukemia or any other type of neoplastic diseases in test animals, although other effects, including pancytopenia, have been widely observed.[1]

E.2.2 Benzene Exposure Limits

It should be noted that where the health effects described above have been associated with benzene exposure, the exposure has been at occupational levels. That is, the benzene exposure levels associated with the effects have been high (10 ppm up to hundreds of parts per million of benzene, except in a few cases of exposures to 2 to 3 ppm benzene) or they have been unknown.

Benzene exposure was first associated with health effects in occupational settings, so initial attempts to limit benzene exposures were aimed at occupational exposures. With recognition of the toxic effects of benzene and its greatly expanded use after 1920, several occupational exposure limits were established in the United States.[3] These limits, originally in the range of 75 to 100 ppm, were successively lowered as more information on benzene toxicity became known.

For example, the American Conference of Governmental Industrial Hygienists (ACGIH) recommended a benzene threshold limit value of 100 ppm in 1946, 50 ppm in 1947, 35 ppm in 1948, 25 ppm in 1949, and 10 ppm in 1977.[3,4] The National Institute for Occupational Safety and Health recommended an exposure limit of 10 ppm in 1974 and revised it downward to 1 ppm in 1976.[5] The current Occupational Safety and Health Administration permissible exposure limit is 10 ppm[6] (a lower limit of 1 ppm is currently in litigation*).

Occupational exposure limits were initially established to protect workers from adverse changes in the blood and blood-forming tissues. The most recently recommended or pending limits of 1 ppm and 10 ppm are based on the conclusion that benzene is leukemogenic in man (NIOSH[5] and OSHA[7]) or a suspected carcinogen in man (ACGIH[4]).

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*A benzene standard with a limit of 1 ppm was proposed by OSHA May 27, 1977, (42 FR 27452) and promulgated February 10, 1978 (43 FR 5918). This standard was struck down October 5, 1978, by the U.S. Fifth Circuit Court of Appeals. The U.S. Department of Labor appealed the decision, and the Supreme Court agreed to hear arguments on the case during its fall 1979 term.

E.2.3 Health Effects at Environmental Exposure Levels

Little information is available on health effects of nonoccupational exposures of the general populace to benzene. Virtually all of the studies cited[1,2] were on the working population (mostly males) exposed to higher than ambient benzene levels on a work cycle. Applying these studies to chronic (24 hours per day) low-level exposure to the general population (including infants, the ill, and the elderly) requires extrapolation.

The recent analysis of benzene health effects[1] concluded that the evidence of increased risk of leukemia in humans on exposure to benzene for various time periods and concentrations was overwhelming but that data were not adequate for deriving a dose-response curve.

However, EPA's Carcinogen Assessment Group (CAG), acknowledging the absence of a clear dose-response relationship, has estimated the risk of leukemia in the general population from low-level benzene exposure.[8] Data from three epidemiological studies of leukemia in workers (mostly adult white males) were used to estimate the risk of developing leukemia. A no-threshold linear model was used to extrapolate this estimated risk to the low levels (below 5 to 10 ppb) to which some populations may be exposed.

The annual risk factor derived for benzene-induced leukemia was 0.34 case per 10^6 ppb person-years. For example, if 3 million persons are chronically exposed to 1 ppb benzene, the model predicts there will be 1.02 cases of leukemia (3 × 0.34) per year in that population. Use of a "linear" model means that the model would predict the same number of leukemia deaths among 3 million people exposed to 1 ppb benzene as among 1 million people exposed to 3 ppb.

The risk factor (0.34 case per 10^6 ppb person-years) was used in estimating the number of leukemia deaths as attributable to benzene emissions from maleic anhydride plants. Other effects of benzene exposure (including fatalities from causes other than leukemia) were not included in the estimated number of deaths. The risk factor equated one leukemia case to one death (that is, each case was presumed fatal).

Several sources of uncertainty are present in the risk factor. First, the retrospective occupational exposure estimates may be inaccurate. CAG calculated the 95-percent confidence intervals for this risk factor to be 0.17 to 0.66 if exposure estimates in the three studies extrapolated are precisely correct, and 0.13 to 0.90 if exposure estimates are off by a factor of 2. Second, the composition of the exposed populations around maleic anhydride plants may vary from that of the populations used as a basis for the CAG estimate. Third, the true dose-response relationship for benzene exposure may not be a linear no-threshold relationship at the low concentrations to which the general population may be exposed. Fourth, the risk factor includes only leukemia deaths and not other health risks. No quantitative estimate of the error in the risk factor due to the latter two uncertainties has been attempted.

E.3 POPULATION DENSITY AROUND MALEIC ANHYDRIDE PLANTS

The population groups "at risk" to benzene exposure (those residing around maleic anhydride plants) were determined[9] from the 1970 Bureau of the Census Master Enumeration District List (MED List) for the area surrounding each plant site.

Each plant site was located by latitude and longitude on a grid system having grids of 10 sq km. The population was determined from the MED List for each grid block within 20 km of each plant site. There were thus approximately 125 population grids for each plant site.

Circles of radii 0.1, 0.3, 0.5, 0.7, 1.0, 1.5, 2.0, 10, and 20 km were overlaid on each grid, and the population within each annular (ring or doughnut-shaped) area was determined. Where grid blocks overlapped two annular areas, the population was assumed to be uniformly distributed and was assigned proportionately to each area. The population in each annular area was considered to be the population exposed to the estimated benzene concentration at the midpoint of that area. The estimated total populations exposed as a function of distance from the plant site are reported in Table E-1.

The method used contains potential sources of error. First, the assumption of uniform population distribution within grid blocks and annular areas may not be correct. For urban areas the assumption is probably valid, but it may introduce some error for rural areas (which is the case with one plant) 10 to 20 km from the site. Another source of error is the use of 1970 population data. However, these are the latest available in the form required. The contractor deriving the population figures made no numerical estimates of probable errors.

TABLE E-1. TOTAL POPULATION EXPOSED AS A FUNCTION OF DISTANCE FROM PLANT SITE

	Population at various distances from plant (km)								
Plant	0.1-0.3	0.3-0.5	0.5-0.7	0.7-1.0	1.0-1.5	1.5-2.0	2.0-10.0	10.0-20.0	Total population
Monsanto	1,300	2,300	3,500	7,400	18,200	22,400	747,000	784,000	1,586,100
Reichhold (Ill.)	0	0	100	300	700	600	11,700	17,600	31,000
U.S. Steel	300	500	700	1,500	3,900	5,000	294,000	873,000	1,178,900
Reichhold (N.J.)	0	0	0	200	800	3,800	651,000	2,358,000	3,013,800
Tenneco	200	400	600	1,300	3,500	7,400	310,000	1,001,000	1,324,400
DENKA	500	1,000	1,500	3,100	7,600	10,800	481,000	739,000	1,244,500
Ashland	0	0	0	0	0	0	19,800	140,000	159,800
Koppers	0	150	200	500	1,400	3,700	297,000	810,000	1,112,950
Total population	2,300	4,350	6,600	14,300	36,100	53,700	2,811,500	6,722,600	9,651,450
Average population density (persons/sq km)	1,150	1,090	1,100	1,120	1,150	1,220	1,170	890	960

SOURCE: Letter Report from Energy and Environmental Analysis, Inc., to U.S. Environmental Protection Agency.

E.4 POPULATION EXPOSURES, RISKS, AND MORTALITIES

E.4.1 Summary of Methodology

The methodology for estimating leukemia deaths due to exposure to benzene emissions from maleic anhydride plants follows.

A typical (or "model") maleic anhydride plant was developed, based on a nominal maleic anhydride production capacity of 22,700 Mg/yr (metric tons per year). Its benzene emissions were estimated to be 190 kg/hr from the product recovery absorber, based on a 94.5-percent benzene conversion rate in the reactor and no benzene emission controls. An additional 2.6 kg/hr benzene was estimated to be emitted from storage, handling, and fugitive sources.[10,11]

The Industrial Source Complex Dispersion (ISCD) model, urban mode 2, was used to estimate mean annual benzene concentrations out to 20 km from the model plant.[12] Pittsburgh meteorological data were used in the dispersion model.

Dispersed benzene concentrations from the model plant were corrected to reflect actual plant capacity and degree of benzene emission controls currently exercised in each plant.[9,12]

The population (1970) around each actual plant location was correlated with corrected benzene concentrations to yield benzene dose in 10^6 ppb person-years per year. The methods for determining populations[9] are described in Part E.3 of this appendix.

From health effects data, the EPA Carcinogen Assessment Group[8] derived a leukemia risk estimate of 0.34 death per 10^6 ppb person-years from exposure to benzene. The methodology for estimating the leukemia risk factor[8] is described in Part E.2.3 of this appendix.

The leukemia deaths per year attributable to exposure to benzene from maleic anhydride plants were estimated by multiplying 3.4×10^{-7} ppb person-years exposure times the exposure in "ppb person-years per year."

Leukemia deaths were estimated for benzene exposures from existing maleic anhydride plants, assuming no control of fugitive, storage, or handling emissions, but four different degrees of benzene emission control on the product recovery absorbers. These four conditions, termed "control alternatives," are:

A. No regulation by any standard; that is, the current conversion rate (94.5 percent) and the current state of benzene emission control on the product recovery absorber in each plant (varies from no control to 90, 97, or 99 percent control);

B. 97 percent reduction in benzene emissions;

C. 99 percent reduction in benzene emissions by thermal incineration; and

D. 99 percent reduction in benzene emissions by carbon adsorption.

For Alternatives B, C, and D, a benzene conversion rate of 90 percent was assumed, resulting in an estimated uncontrolled benzene emission rate of 345.4 kg/hr from the product recovery absorber of the model plant.

For all control alternatives, two control system failure scenarios were assumed. First, it was assumed that benzene control systems on the product recovery absorbers would fail for 3 hours 15 times per year and would be limited to emissions of 250 kg of benzene per plant (on the basis of the model plant capacity) during each failure. The uncontrolled benzene emission rate from the model plant product recovery absorber is 190 kg/hr, so the 250-kg emission 15 times per year would be equivalent to 20 hr/yr of uncontrolled emissions. The second failure condition assumed that benzene control systems would fail for 48 hr/yr, during which product recovery absorber emissions would be 190 kg/hr (based on the model plant) for the full 48 hours.

It was assumed further that the atmospheric dispersion of these unanticipated emissions would follow the same pattern as during normal (controlled) operations and that emissions from fugitive and storage sources would be unaffected by the failures.

E.4.2 Estimates of Leukemia Deaths

The general equation for estimating the number of leukemia deaths attributable to benzene emissions from a particular plant (e.g., plant X) under either normal or uncontrolled operations is:

$$D_x = \sum_{i = 0.1\text{-}0.3}^{10\text{-}20} (R)(P_i/10^6)(B_i)(C/22.7)(f), \qquad (1)$$

in which

D_x = estimated number of leukemia deaths per year from benzene emissions from the plant (e.g., plant X).

R = the risk factor (0.34 death per 10^6 ppb person-years).

P_i = population at risk, in area (i) around plant X (Table E-1).

B_i = mean annual average benzene concentration (ppb) in area (i) around plant X for the control alternative selected (A,B,C, or D) or for the uncontrolled condition (Table E-3).[12]

C = capacity of plant X (in thousands of metric tons of maleic anhydride produced per year; the capacity of the model plant is 22,700 metric tons/yr).

f = fraction of plant X output produced from benzene (e.g., if 20 percent of plant capacity is from n-butane feedstock, f = 0.80).

i = the particular area in which P_i and B_i occur (i progresses from the area 0.1 to 0.3 km from the plant to the area 10 to 20 km from the plant).

Σ = summation of deaths from all areas (i).

Values for P_i are given in Table E-1. Values for R, C, f, current control levels, and control levels under failure conditions are given in Table E-2. Values of B_i for all control levels of interest, as determined from Table 32 of Reference 12, are given in Table E-3.

Leukemia deaths per year were estimated for each plant for both controlled and uncontrolled conditions, using Equation 1. The total estimated number of leukemia deaths per year attributable to benzene emissions from all plants was determined for each condition by the equation:

$$\text{Total estimated number of leukemia deaths/yr } (D_t) = D_1 + D_2 + \dots + D_8 \,. \tag{2}$$

The total numbers of estimated leukemia deaths attributable to benzene emissions from maleic anhydride plants are given in Table E-4 on a plant-by-plant basis, in deaths per year, for three control alternatives (current, 97 percent, and 99 percent), the uncontrolled condition (0 percent), and the shutdown condition (storage, handling, and fugitive emissions only). For purposes of Table E-4, Alternatives C and D are indistinguishable and have been combined. The alternatives defined in Table E-4 thus differ slightly from the nomenclature previously used (A,B,C, and D).

TABLE E-2. FACTORS IN ESTIMATING LEUKEMIA DEATHS FROM BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS

Plant	Capacity, 10^3 metric tons/yr (C)	Fraction from benzene (f)	Benzene control level at product recovery absorber (%)		Risk factor (R)
			Current level	Under control system failure conditions	
Monsanto	47.6	0.8	0	0	0.34
Reichhold (Ill.)	20	1.0	90	0	0.34
U.S. Steel	38.5	1.0	90	0	0.34
Reichhold (N.J.)	13.6	1.0	97	0	0.34
Tenneco	11.8	1.0	0	0	0.34
DENKA	22.7	1.0	97	0	0.34
Ashland	27	1.0	0	0	0.34
Koppers	15.4	1.0	99	0	0.34

TABLE E-3. ESTIMATED MEAN ANNUAL AVERAGE BENZENE CONCENTRATIONS (B_i) AROUND MODEL MALEIC ANHYDRIDE PLANTS[a]

Column (1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)	(9)
	Benzene concentration (ppb)							
		Unregulated plant (94.5% conversion) with PRA[b] emissions reduced by			Regulated plant (90% conversion) with PRA[b] emissions reduced by			
Distance from plant (km)	0%[c]	90%	97%	99%	97%	99% (incineration)	99% (adsorption)	100%[d]
0.1-0.3	5.27	2.3	2.0	1.95	2.11	1.99	2.12	1.9
0.3-0.5	6.12	1.37	0.97	0.87	1.1	0.87	0.99	0.82
0.5-0.7	5.7	0.96	0.6	0.49	0.73	0.53	0.61	0.43
0.7-1.0	4.8	0.7	0.38	0.29	0.51	0.33	0.38	0.245
1.0-1.5	3.6	0.5	0.24	0.17	0.33	0.20	0.22	0.13
1.5-2.0	2.55	0.34	0.155	0.105	0.22	0.13	0.14	0.076
2.0-10.0	0.49	0.057	0.025	0.0155	0.037	0.023	0.021	0.0103
10.0-20.0	0.135	0.0155	0.0062	0.0038	0.0097	0.0057	0.0054	0.0025

[a] Values in Columns 2, 3, 4, 5, and 9 were derived by adding storage, handling, and fugitive source benzene concentrations from Table 32 of Reference 12 to 100 percent, 10 percent, 3 percent, 1 percent, or 0 percent of the "Alternative A" product recovery absorber (PRA) benzene concentrations, respectively, for each distance. Resulting concentrations were plotted vs. distance on log-log paper, and the concentration at the midpoint of each segment was determined. Micrograms/m^3 were converted to ppb by dividing by 3.2. Values in Columns 6, 7, and 8 were derived by adding the storage, handling, and fugitive source benzene concentrations to the PRA concentrations for the appropriate control alternative (B, C, or D) from Table 32, except that: (1) for all three options, PRA concentrations were increased by (1 - 0.9)/(1 - 0.945), or 1.82X, to account for higher emissions due to changing the benzene conversion rate from 94.5 percent to 90 percent; and (2) for Alternatives C and D (Columns 7 and 8), PRA concentrations were further increased by (1 - 0.99)/(1 - 0.995), or 2X, to account for higher emissions due to changing the required control system efficiency from 99.5 percent to 99 percent. Total concentrations were plotted vs. distance, on log-log paper, and the concentration at the midpoint of each segment was determined and recorded in Table E-3.

[b] Product recovery absorber.

[c] Condition of no control or control system failure.

[d] Fugitive, storage, and handling emissions only.

SOURCE: H. E. Cramer Co., Inc. Dispersion Model Analysis of the Air Quality Impact of Benzene Emissions from a Maleic Anhydride Plant for Four Emission Control Options. EPA Contract Number 68-02-2507. August 1978.

TABLE E-4. ESTIMATED LEUKEMIA DEATHS DUE TO BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS

Column (1)	(2)	(3)	(4)	(5)	(6)	(7)
		Estimated leukemia deaths/yr with product recovery absorber benzene emissions reduced by				
		Current control levels		97% by standard	99% by standard	
Plant	0 percent (control system failure condition)	% current control	With no control failures or shutdown	With no control failures or shutdown	With no control failures or shutdown	100%[a] (plant shutdown condition)
Monsanto	0.3827	0	0.3827	0.033	0.0212	0.0122
Reichhold (Ill.)	0.00425	90	0.00054	0.000358	0.000222	0.000125
U.S. Steel	0.1756	90	0.0214	0.0140	0.00869	0.00447
Reichhold (N.J.)	0.1326	97	0.00647	0.00647[b]	0.00593	0.002657
Tenneco	0.0586	0	0.0586	0.00458	0.00283	0.00141
DENKA	0.1437	97	0.0082	0.0082[b]	0.00744	0.00400
Ashland	0.0116	0	0.0116	0.000845	0.000507	0.000224
Koppers	0.0632	99	0.00200	0.00200[b]	0.00200[b]	0.001356
Total deaths/yr[c]	0.9723	-	0.492	0.0695	0.0488	0.0264

[a] Fugitive, storage, and handling emissions only.

[b] Where current control levels result in fewer estimated deaths than with the control levels required by the standard, the lesser figure is used.

[c] These figures are for each condition if it were to occur 8,760 hr/yr. Figures in Table E-5 assume 8,000 hr/yr of plant operations and so are lower.

The death estimates shown in Table E-4 are for each condition if it were to occur continuously, and so were weighted in accordance with the following operating assumptions. Product recovery absorbers (PRA) were assumed to operate 8,000 hr/yr. Storage, handling, and fugitive emissions were assumed to occur 8,760 hr/yr; that is, they continued even during plant shutdown. During the first failure condition, with 15 control system failures of 3 hours each, the period of normal control system operations would be 7,955 hr/yr. During failure, emissions would be equivalent to 20 hr/yr of uncontrolled operations, based on the model plant and a 94.5-percent benzene conversion rate. During the second failure condition, with 48 hours of control system failure, the period of normal control system operations would be 7,952 hr/yr.

The risk factor ($0.34/10^6$ ppb person-years) is based on continuous exposure (8,760 hr/yr).[8] The estimated leukemia deaths per year were therefore determined by weighting deaths from Table E-4 in accordance with the duration of each condition. For the first failure condition, this would be: controlled operations (7,955 hr/yr), uncontrolled operations (equivalent of 20 hr/yr), and storage, handling, and fugitive emissions only (an additional 785 hr/yr). Thus, the total estimated number of leukemia deaths per year was determined as:

$$\begin{aligned} D_T = {} & [(D_t \text{ under no-control conditions}) \times (20/8{,}760)] \\ & + [(D_t \text{ under current or regulated condition;} \\ & \text{current level, 97, or 99\%}) \times (7{,}955/8{,}760)] \\ & + [(D_t \text{ under shutdown conditions}) \times (785/8{,}760)] \, . \end{aligned} \tag{3a}$$

The equivalent of 20 hours of uncontrolled emissions instead of the 45-hour actually assumed emission period is used to put the emission rate into the terms used in the dispersion model report in order to simplify the calculations. However, fugitive, storage, and handling emissions continue for the remaining 25 hours of this period as well as the 760 hr/yr when the plant is shut down, for a total of 785 hr/yr.

Similarly, for the second failure condition, leukemia deaths per year were determined as:

$$\begin{aligned} D_T = {} & [(D_t \text{ under no-control conditions}) \times (48/8{,}760)] \\ & + [(D_t \text{ under current or regulated condition;} \\ & \text{current level, 97, or 99\%}) \times (7{,}952/8{,}760)] \\ & + [(D_t \text{ under shutdown conditions}) \times (760/8{,}760)] \ . \end{aligned} \tag{3b}$$

Deaths per year resulting from operations based on 8,000 hr/yr of product recovery absorber operations with no control failures and an additional 760 hr/yr of only storage, handling, and fugitive emissions were determined by:

$$\begin{aligned} D_T = {} & [(D_t \text{ under current or regulated condition;} \\ & \text{current level, 97, or 99\%}) \times (8{,}000/8{,}760)] \\ & + [(D_t \text{ under shutdown conditions}) \times (760/8{,}760)] \ . \end{aligned} \tag{3c}$$

The total estimated leukemia deaths per year for the control options of interest are summarized in Table E-5. It is apparent that the brief control system failure conditions have little effect on annual exposure and deaths. Also, even with continuous 99 percent control of product recovery absorber benzene emissions, there will still be an estimated 0.0488 leukemia death per year. From Table E-4 (last two columns), it can be seen that 0.0264 (or 54 percent) of these results from fugitive, handling, and storage emissions, and 0.0224 death (46 percent) from residual product recovery absorber emissions.

E.4.3 Example of Leukemia Death Calculation

Values in Table E-4 were determined in accordance with the following example, using the U.S. Steel plant. From Table E-2, R = 0.34, C = 38.5, and f = 1 for all distances, so Equation 1 may be simplified to:

$$D_x = (R)(C/22.7)(f)(1/10^6) \sum_{i = 0.1\text{-}0.3}^{10\text{-}20} P_i B_i \ , \tag{4}$$

or

$$D_x = 5.77 \times 10^{-7} \sum_{i = 0.1\text{-}0.3}^{10\text{-}20} P_i B_i \ .$$

P_i values for each area (i) around this plant are taken from Table E-1. Table E-2 shows that this plant has a current control level of 90 percent, so B_i values for 90 percent control in each area (i) are taken from Table E-3. $\Sigma P_i B_i$ is calculated as follows:

TABLE E-5. ESTIMATED LEUKEMIA DEATHS FROM BENZENE EMISSIONS FROM MALEIC ANHYDRIDE PLANTS[a]

	Estimated leukemia deaths		
Product recovery absorber control alternative	With no control failures[b]	With 15 control failures/yr[c]	With 48 hr/yr of control failures[d]
No regulation by EPA (current level of control)--Alternative A	0.45	0.45	0.45
97% benzene removal (by incineration or adsorption)--Alternative B	0.066	0.068	0.071
99% benzene removal (by incineration or carbon adsorption)--Alternatives C and D	0.047	0.049	0.052

[a] It should be recognized that considerable uncertainty is associated with the cases of leukemia that appear in the table. First, the cases were calculated based on an extrapolation of leukemia risk associated with a healthy white male cohort of workers to the general population, which includes men, women, children, infants, the aged, nonwhites, and the unhealthy. Second, there are potential errors in estimating the benzene levels to which people in the vicinity of maleic anhydride plants are exposed. Also, the number of cases includes consideration of only one effect of benzene; i.e., leukemia. Benzene has also been indicated to cause aplastic anemia, cytopenias, and the development of chromosomal aberrations. In addition, the benefits to the general population of controlling other hydrocarbon emissions from maleic anhydride manufacture are not quantified.

For comparison, 665.95 total leukemia deaths per year would be expected in the exposed population of 9.65 million residing within 20 km of the eight existing maleic anhydride plants. This figure is based on the overall U.S. leukemia death rate of 6.9 per population of 100,000 for 1972, 1973, and 1975.[13] The rate ranged from 6.7 to 7.6 among the six States where maleic anhydride plants are located.

CAG has estimated that there are 90 leukemia deaths per year nationwide because of benzene exposure.[9]

[b] Calculation described in Equation 3c.

[c] Calculation described in Equation 3a.

[d] Calculation described in Equation 3b.

Area (i)	P_i (from Table E-1)	B_i (from Table E-3) for 90% control	$P_i \times B_i$
0.1 - 0.3	300	2.3	690
0.3 - 0.5	500	1.37	685
0.5 - 0.7	700	0.96	672
0.7 - 1.0	1,500	0.7	1,050
1.0 - 1.5	3,900	0.5	1,950
1.5 - 2.0	5,000	0.34	1,700
2.0 - 10.0	294,000	0.057	16,758
10.0 - 20.0	873,000	0.0155	13,532

$\Sigma P_i B_i = 37{,}037$

$$D_x = 5.77 \times 10^{-7} (37{,}037) \quad = \quad 0.0214.$$

This value (0.0214) is shown for U.S. Steel in Column 4 of Table E-4, assuming no control system failures. The same procedure is used to determine deaths assuming 100 percent control failure, by using B_i values for 0 percent control from Table E-3. The resultant value (0.1756) is in Column 2 of Table E-4. All values in Table E-4 were determined with Equation 1 in a like manner, using B_i values for the appropriate control level and R, C, f, and P_i values for the specific plant. Total deaths were obtained by summing the deaths attributable to each plant, using Equation 2. Totals are shown in the lower section of Table E-4.

Deaths listed in Table E-5 were determined from Equations 3a, 3b, and 3c. Deaths at, say, 97 percent control, with 15 control failures per year were determined from Equation 3a as:

$$D_T = 0.0695\ (7{,}955/8{,}760) + 0.9723\ (20/8{,}760) + 0.0264\ (785/8{,}760) = 0.068.$$

With no control failures, deaths per year were determined from Equation 3c as:

$$D_T = 0.0695\ (8{,}000/8{,}760) + 0.0264\ (760/8{,}760) = 0.066.$$

Other deaths listed in Table E-5 were determined in the same manner.

E.4.4 Estimate of Leukemia Risk

The estimated leukemia deaths shown in Tables E-4 and E-5 are based on estimates of mean annual average benzene concentrations around maleic anhydride plants. Because atmospheric dispersion patterns are not uniform, some population groups will receive above-average benzene exposures and will therefore incur a higher risk (or probability) of contracting leukemia.

Maximum annual risk is the estimated probability to someone, who is constantly exposed to the highest maximum annual average benzene concentration in the ambient air around a particular source for a year, of contracting leukemia because of exposure to benzene emissions from that source. Maximum lifetime risk can be found by multiplying the maximum annual risk by 70 years.

Maximum annual risks of leukemia deaths in exposed population groups were estimated from the maximum annual average benzene concentrations estimated[12] for various distances from the model plant, using the equation:

Maximum annual risk of leukemia death for an individual/yr
= (B_i, the max annual benzene concentration, ppb)
× (R, the risk factor of 0.34 death per 10^6 ppb person-years)
× (10^{-6}, which is one person as a fraction of 1 million), or

Maximum annual risk of leukemia death for an individual/yr

$$= (3.4 \times 10^{-7})(\text{max } B_i \text{ in ppb}) \ , \tag{5a}$$

or

$$= (1.06 \times 10^{-7})(\text{max } B_i \text{ in } \mu g/m^3) \ . \tag{5b}$$

For the uncontrolled model plant, the estimated maximum annual average benzene concentration (max B_i) is 35.7 $\mu g/m^3$ at 0.3 km from the plant (Table 14 of Reference 12). The estimated maximum annual risk or probability of leukemia death for an individual is 3.8×10^{-6} ($1.06 \times 10^{-7} \times 35.7$) per year.

For the same plant with 97 percent control of benzene emissions from the product recovery absorber, the maximum estimated risk is 2.2×10^{-6} per person per year, which occurs 0.1 km from the plant (Table 22 of Reference 12). This calculation is shown in Figure E-1. Calculations for maximum risk are summarized in Table E-6.

Maximum annual risks of leukemia from product recovery absorber emissions only were calculated in a similar manner, using maximum annual average concentrations from Table 9 of Reference 12. Calculations are shown in Table E-7. The maximum risk associated with the uncontrolled model plant is 3.4×10^{-6} per person per year. With 97 percent benzene control, the

$$\underset{(\text{in } \mu g/m^3)}{\text{Max } B_i} = \underset{\substack{\text{Max } B_i\\ \text{due to}\\ \text{storage}}}{9.09} + \underset{\substack{\text{Max } B_i\\ \text{due to}\\ \text{handling}}}{2.77} + \underset{\substack{\text{Max } B_i\\ \text{due to}\\ \text{fugitive}\\ \text{sources}}}{8.83} + \Bigg[\underset{\substack{\text{Max } B_i \text{ due to product}\\ \text{recovery absorber (PRA)}}}{0.0317} \times \underset{\substack{\text{Increase in}\\ B_i \text{ due to}\\ \text{changing}\\ \text{conversion}\\ \text{rate from}\\ 94.5\% \text{ to}\\ 90\%}}{1.82}$$

$$\times \underset{\substack{\text{Change in}\\ \text{control}\\ \text{system}\\ \text{efficiency}\\ \text{from}\\ 99.5\% \text{ to}\\ 99\%}}{2} \times \underset{\substack{\text{Percent of}\\ \text{total annual}\\ \text{hours}}}{\frac{7{,}955}{8{,}760}}\Bigg] + \Bigg[\underset{\substack{\text{Max } B_i \text{ at}\\ 0.1 \text{ km due}\\ \text{to PRA during}\\ \text{control system}\\ \text{malfunction}}}{0.497} \times \underset{\substack{\text{Percent of total}\\ \text{hours that}\\ \text{occur during}\\ \text{control system}\\ \text{malfunction}}}{\frac{20}{8{,}760}}\Bigg] = 20.796\ \mu g/m^3$$

$$\text{Maximum Risk} = (1.06 \times 10^{-7})(20.796) = 2.2 \times 10^{-6}.$$

Figure E-1. Example calculation of maximum annual risk for 97 percent control.

TABLE E-6. EFFECT OF MODIFYING EMISSION CONTROL PARAMETERS ON ESTIMATED MAXIMUM LEUKEMIA RISK

Factors Column (1)	Emissions at point of maximum risk (in μg/m^3) (2)	(3)	(4)	(5)	(6)	(7)
Benzene conversion rate (%)	94.5%	94.5%	94.5%	90%	90%	90%
Control method[a]	either	inc.	ads.	either	inc.	ads.
Control level (%)	97	99.5	99.5	97	99	99
15 failures/yr with excess emissions	NO	NO	NO	NO	NO	NO
48 hr/yr of failure with excess emissions	NO	NO	NO	NO	NO	NO
Benzene from storage, handling, and fugitive sources[b]	20.69	20.69	20.69	20.69	20.69	20.69
Benzene from product recovery absorber control system:	0.0317[b]	0.00405[b]	0.0395[b]	0.0576	0.01473	0.1436
Total benzene concentration	20.7217[b]	20.6941[b]	20.7295[b]	20.7476	20.7047	20.8336
Risk:[c] probability of leukemia death per person per year	2.2×10^{-6}	2.2×10^{-6}	2.2×10^{-6}	2.2×10^{-6}	2.2×10^{-6}	2.2×10^{-6}

[a] Inc. = incineration; ads. = adsorption; either = either method.

[b] Benzene concentrations (μg/m^3) taken directly from Reference 12.

[c] Risk = (B_i, μg/m^3) × (0.34, the risk per 10^6 ppb person years/yr) × (10^{-6}, which is one person as a fraction of 10^6) × (1/3.2, to convert μg/m^3 to parts per billion). Risk = $0.106 \times 10^{-6} \times B_i$. Decimal places are retained only to show effects of parameters changes. For comparison, the maximum risk associated with emissions from the unregulated model plant is 3.8×10^{-6} per person per year.

TABLE E-7. CALCULATIONS FOR MAXIMUM ANNUAL LEUKEMIA RISK FROM MODEL PLANT PRODUCT RECOVERY ABSORBER EMISSIONS ONLY[a]

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1. UNCONTROLLED $(32.0)^{b} \times (1.06 \times 10^{-7})^{a} = 3.4 \times 10^{-6}$

2. 97% CONTROL (e.g., incineration)

$$\frac{8{,}000}{8{,}760} \times (1.91)^{c} \times (1.06 \times 10^{-7}) = 1.8 \times 10^{-7}$$

3. 99% CONTROL (e.g., incineration)

$$\frac{8{,}000}{8{,}760} \times (0.466)^{c} \times (1.06 \times 10^{-7}) = 4.5 \times 10^{-8}$$

---

[a] Risk = (B_i, μg/m³) × (0.34, the risk per 10^6 ppb person-years/yr) × (10^{-6}, which is one person as a fraction of 10^6) × (1/3.2, to convert μg/m³ to ppb). Risk = $1.06 \times 10^{-7} \times B_i$. Decimal places are retained only to show effects of parameter changes.

[b] Table 9, Cramer Report (Reference 12), in μg/m³.

[c] These numbers represent the respective ambient concentrations, given in Table 9 of the Cramer report. The Cramer report ambient concentrations are multiplied by 1.82 because of the increase in allowable emissions under the standard (based on a 90-percent benzene conversion rate) over emissions in the Cramer model based on 94.5 percent benzene conversion. For 99 percent control, the revised concentrations are then multiplied by 2 because the Cramer report is based on 99.5 percent control, while the control level considered is 99 percent.

maximum risk is 1.8×10^{-7} per person per year. For 99 percent control, the maximum risk is 4.5×10^{-8} per person per year.

E.4.5 Validity of Estimates

Several potential sources of error exist in the estimated factors used in Equation 1 (R, P_i, B_i, C, f). Possible errors in the risk factor (R) are discussed in Part E.2.3, and sources of error in populations "at risk" (P_i) are discussed in Part E.3. The validity of the other factors is discussed below. Readers should note that the number of significant figures carried in the decimals in Tables E-3 through E-7 is not an indication of their accuracy but rather a means of enabling readers to duplicate the calculations and to visualize the magnitude of difference among the options considered.

E.4.5.1 Plant capacity (C, f). Plant capacities (C) in thousands of metric tons per year were estimated by projecting output to 1982 and assuming operation at full capacity. The factor (f) represents the fraction of total plant output produced using benzene as the feedstock. Estimates of (f) are based on current plant operations.

Because plant capacities are often nominal values, the capacity and production estimates may be inaccurate either on the high or low side, affecting (C). Plants may start using n-butane as a feedstock to a greater or lesser degree, affecting (f). The degree of error in the estimates of (C) and (f) cannot be defined in numerical terms.

E.4.5.2 Benzene Concentrations (B_i)

The estimated benzene concentrations are derived from several factors, as follows:

- Configuration of the model plant,
- Emission rates from the model plant, and
- Dispersion patterns of the emissions.

Numerical error limits could not be calculated for these factors, but their qualitative effects on the estimated number of leukemia deaths are discussed below.

The configuration of the model plant assumes a given area (4,500 sq m), with storage tanks and product recovery absorbers at specific locations and heights, and with fugitive sources uniformly distributed at the plant site. Emissions from the model plant have been estimated from several sources and uniform emission rates assumed. Current levels of control on the existing recovery absorbers were determined for each plant in calculating emissions, and these control levels were assumed to be the same for future operations. When the model plant was applied to existing plant emissions, it was assumed that benzene emissions varied in direct proportion to plant capacity.

It is unlikely that any plant duplicates the model plant precisely, and it is recognized that the estimated benzene emissions will be in error to some degree. For example, the benzene emissions used in the original calculations (190 kg/hr from the product recovery absorber and 1.8 kg/hr from storage and handling)[11] were estimated to have 95 percent confidence limits of ±58 percent and ±20 percent, respectively. Confidence limits were not determined for fugitive emissions (0.8 kg/hr).[10] No numerical estimates on the other potential errors could be determined, nor could an overall error range be estimated from all factors.

Additional sources of potential error occur in the atmospheric dispersion model for benzene emissions. First, the model used weather data from the Greater Pittsburgh Airport to project dispersion patterns for all existing plants. Pittsburgh has a high frequency of west winds, so the model may tend to overstate maximum risks for plants in other areas. Because the model (the Industrial Source Complex Dispersion Model, urban mode 2) is intended to account for effects of urban roughness and heat sources, it may introduce inaccuracies when applied to less urbanized areas. Second, the model assumes there is no loss of benzene from atmospheric reactions or ground level absorption. If such losses occur, the actual concentration of benzene will be less than the estimated values. Third, atmospheric dispersion patterns of benzene emitted during control failures may not be the same as those over the course of a year. However, any such deviation from average would have little effect on total death estimates because of the short failure periods assumed. This can be seen in Table E-4. It is estimated that benzene concentrations predicted by the dispersion model may be inaccurate by a factor of 2.[12]

A final source of error is that the model measures benzene dispersion only to 20 km. If the linear risk model is accurate, exposures at distances greater than 20 km, however small, may be important. If such exposures occur, the estimated number of deaths will be higher than estimated here, particularly for the unregulated plants.

E.5 REFERENCES

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